5. TECHNOLOGY CASE STUDIES

A series of case studies have been performed on the three conversion routes for combined heat and power (CHP) applications of biomass—direct combustion, gasification, and cofiring. The studies are based on technology characterizations developed by NREL and EPRI [1], and the technology descriptions are excerpted from that study. Variables investigated include plant size and feed cost, and both cost of electricity and cost of steam are estimated using a discounted cash flow analysis. The economic basis for cost estimates is given below.

Table 5.1: Economic Assumptions

Parameter	Value	Discussion
Basis Year	3rd Qtr, 2001	
Cost Index	Marshall & Swift	
Scale Factor	0.7	
Debt	50%	
Inflation	None	
Capital	7% for 20 years	
Discount Rate	20%	
Salvage Value	0	
Taxes		
Federal	35%	
State	5%	
State Wholesale Excise	0	
Federal Alt. Min. Tax	Not estimated	
Industrial Electricity Purchase Price	3.8 cents/kWh	Chemicals Industry Costs: EIA Manufacturing Consumption of Energy 1998, Table N8.3, corrected to 2001 $
Industrial Steam Purchase Price	$3.3/1000lb	Chemicals Industry Costs: EIA Manufacturing Consumption of Energy 1998, Table N8.3, corrected to 2001 $
Construction Period	2 years	Comparable to EPRI/DOE Technical Characterization
Operating Life	30 years	Comparable to EPRI/DOE Technical Characterization
Stream Factor	90%	
Depreciation	7-year MACRS	For biomass specific operations
	20-year MACRS	For generating equipment, BOP
	5-year MACRS	For biomass qualifying facility sensitivity case
Tax Credit	0.5, 1, 1.5, 2 ¢/kWh	Sensitivity Case
Financial Parameter	NPV(0)	net present value
Feed Costs	-1,0,1,2,3,4 $/MBtu	
Plant Size	25, 50, 75, 100 MW 75 MW 45, 105 MW	Based on electricity only - direct combustion Gasification Cofiring (biomass contribution)

[1] "Renewable Energy Technology Characterizations," EPRI-TR-109469, Electric Power Research Institute, Palo Alto, California, December 1997 (RETC97)

Technology Alternatives

The nearest term low-cost option for the use of biomass is cofiring with coal in existing boilers. Cofiring refers to the practice of introducing biomass as a supplementary energy source in high efficiency boilers. Boiler technologies where cofiring has been practiced, tested, or evaluated, include wall- and tangentially-fired pulverized coal (PC) boilers, cyclone boilers, fluidized-bed boilers, and spreader stokers. The current coal-fired power generating system represents a direct system for carbon mitigation by substituting biomass-based renewable carbon for fossil carbon. Extensive demonstrations and trials have shown that effective substitutions of biomass energy can be made up to about 15% of the total energy input with little more than burner and feed intake system modifications to existing stations. Since large scale power boilers (both utility and independent operators) in the 1999 345 GW (EIA 1999) capacity fleet range from 100 MW to 1.3 GW, the biomass potential in a single boiler ranges from 15 MW to 150 MW. Preparation of biomass for cofiring involves well known and commercial technologies. After tuning the boiler's combustion output, there is little or no loss in total efficiency, implying that the biomass combustion efficiency to electricity would be about 33-37%. Since biomass in general has significantly less sulfur than coal, there is a SO_2 benefit; and early test results suggest that there is also a NO_x reduction potential of up to 20% with woody biomass. Investment levels are very site specific and are affected by the available space for yarding and storing biomass, installation of size reduction and drying facilities, and the nature of the boiler burner modifications. Investments are expected to be in $100 - 700/kW of biomass capacity, with a median in the $180 - 200/kW range.

Another potentially attractive biopower option is based on gasification. Gasification for power production involves the devolatilization and conversion of biomass in an atmosphere of steam or air to produce a medium- or low- calorific gas. This biogas is used as fuel in a combined cycle power generation cycle involving a gas turbine topping cycle and a steam turbine bottoming cycle. A large number of variables influence gasifier design, including gasification medium (oxygen or no oxygen), gasifier operating pressure, and gasifier type. Advanced biomass power systems based on gasification benefit from the substantial investments made in coal-based gasification combined cycle (GCC) systems in the areas of hot gas particulate removal and synthesis gas combustion in gas turbines. They also leverage investments made in the Clean Coal Technology Program (commercial demonstration cleanup and utilization technologies) and in those made as part of DOE's Advanced Turbine Systems (ATS) Program. Biomass gasification systems will also stand ready to provide fuel to fuel cell and hybrid fuel-cell/gas turbine systems, particularly in developing or rural areas without cheap fossil fuels or problematic transmission infrastructure. The first generation of biomass GCC systems would realize efficiencies nearly double that of the existing industry. In a cogeneration application efficiencies could exceed 80%. This technology is very near to commercial availability with mid-size plants operating in Finland, the UK, the Netherlands, and Vermont. Costs of a first-of-a-kind biomass GCC plant are estimated to be in the $1800-2000/kW range with the cost dropping rapidly to the $1400/kW range for a mature plant in the 2010 time frame.

Direct-fired combustion technologies are another option, especially with retrofits of existing facilities to improve process efficiency. Direct combustion involves the oxidation of biomass with excess air, giving hot flue gases that produce steam in the heat exchange sections of boilers. The steam is used to produce electricity in a Rankine cycle. In an electricity-only process, all of the steam is condensed in the turbine cycle while, in CHP operation, a portion of the steam is extracted to provide process heat. Today's biomass-fired steam cycle plants typically use single pass steam turbines. However, in the past decade, efficiency and design features, found previously in large-scale steam turbine generators, have been transferred to smaller capacity units. These designs include multi-pressure, reheat and regenerative steam turbine cycles, as well as supercritical steam turbines. The two common boiler designs used for steam generation with biomass are stationary- and traveling-grate combustors (stokers) and atmospheric fluid-bed combustors. The addition of dryers and incorporation of more-rigorous steam cycles is expected to raise the efficiency of direct combustion systems by about 10% over today's efficiency, and to lower the capital investment from the present $2,000/kW to about $1275/kW.

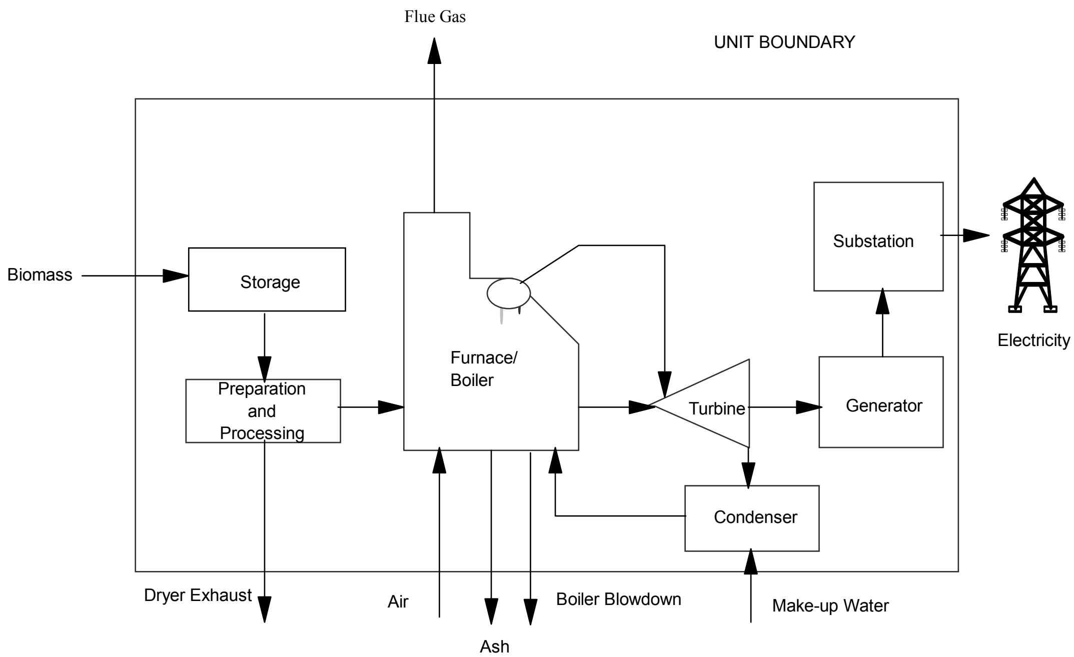
The three technologies are all at either the commercial scale or commercial prototype scale, and have been included in this technology case study. There are additional technologies which are at the conceptual or research and development stage and do not warrant development of a technology case study at this time, but which are potentially attractive from a performance and cost perspective and merit discussion. These technologies include biomass gasification fuel cell processes, and modular systems such as biomass gasification/Stirling engines.

Gasification fuel cell systems hold the promise for high efficiency and low cost at a variety of scales. The benefits may be particularly pronounced at scales previously associated with high cost and low efficiency (i.e., from < 1 MW to 20 MW). Fuel cell based power systems are likely to be particularly suitable for distributed power generation strategies in the U.S. and abroad. Extensive development of molten carbonate fuel cell (MCFC) technology has been conduced under DOE's Fossil Energy Program largely with natural gas as a test fuel. Several demonstration projects are underway in the U.S. for long-term testing of these cells. A limited amount of testing was also done with MCFC technology on synthesis gas from a DOW coal gasifier at DESTEC's facility in Plaquamine, LA. The results from this test were quite promising.

To date, little fuel cell testing has been done with biomass-derived gases despite the several advantages that biomass has over coal in this application. Biomass' primary advantage is its very low sulfur content. Sulfur-containing species is a major concern in fossil fuel-based fuel cell systems because all fuel cells are very sensitive to this contaminant. An additional biomass advantage is its high reactivity. This allows biomass gasifiers to operate at lower temperatures and pressures, while maintaining throughput levels comparable with their fossil fueled counterparts. These relatively mild operating conditions and a high throughput should permit economic construction of gasifiers of a relatively small scale that are compatible with planned fuel cell system sizes. Additionally, the operating temperature and pressure of MCFC units may allow a high degree of thermal integration over the entire gasifier/fuel cell system. Despite these obvious system advantages, it is still necessary for actual test data and market assessments to be obtained to stimulate commercial development and deployment of fuel cell systems.

The Stirling engine is designed to use any heat source, e.g., biomass, and any convenient working gas to generate energy, in this case electricity. The basic components of the Stirling engine include: a compression space and an expansion space with a heater, regenerator, and cooler in between. Heat is supplied to the working gas at a higher temperature by the heater and is rejected at a lower temperature in the cooler. The regenerator provides a means for storing heat deposited by the hot gas in one stage of the cycle and releasing it heat the cool gas in a subsequent stage. Stirling engine systems using biomass are ideal for remote applications, stand alone or cogeneration applications, or as backup power systems. A feasibility test of biomass gasification Stirling engine generation has been performed by Stirling Thermal Motors using a 25 kW engine connected to a small Chiptec updraft gasifier. While the results were encouraging, further demonstration of the concept is required.

Figure 5.1: Direct-fired Biopower Facility



Direct-Fired Biomass

Direct combustoin, illustrated in Figure 5.1, involves the oxidation of biomass with excess air, giving hot flue gases that produce steam in the heat exchange section of boilers. The steam is used to produce electricity in a Rankine cycle; usually, only electricity is produced in a condensing steam cycle, while electricity and steam are cogenerated in an extracting steam cycle. Today's biomass-fired steam cycle plants typically use single-pass steam turbines. However, in the past decade, efficiency and design features, found previously in large scale steam turbine generators (>200 MW), have been transferred to smaller capacity units. These designs include reheat and regenerative steam cycles as well as supercritical steam turbines. The two common boiler designs used for steam generation with biomass are stationary- and traveling-grate combustors (stokers) and atmospheric fluid-bed combustors.

All biomass combustion systems require feedstock storage and handling systems. The 50-MW Burlington, Vermont, McNeil station, which uses a spreader-stoker boiler for steam generation, has a typical feed system for wood chips (Wiltsee and Hughes 1995). Whole tree chips are delivered to the plant gate by either truck or rail. Fuel chips are stored in open piles (about a 30 day supply on about 3.25 ha of land), fed by conveyor belt through an electromagnet and disc screen, then fed to surge bins above the boiler by belt conveyors. From the surge bins the fuel is metered into the boiler’s pneumatic stokers by augers.

Pile burners represent the historic industrial method (Hollenbacher 1992) of wood combustion and typically consist of a two-stage combustion chamber with a separate furnace and boiler located above the secondary combustion chamber. The combustion chamber is separated into a lower pile section for primary combustion and an upper secondary-combustion section. Wood is piled about 3.3 m (10 ft) deep on a grate in the bottom section and combustion air is fed upwards through the grate and inwards from the walls; combustion is completed in a secondary combustion zone using overfire air. Feed is introduced either on top of the pile or through an underfeed arrangement using an auger. The underfeed arrangement gives better combustion control by introducing feed underneath the active combustion zone, but it increases system complexity and lowers its reliability. Ash is removed by isolating the combustion chamber from the furnace and manually dumping the ash from the grate after the ash is cooled. Pile burners typically have low efficiencies (50% to 60%), cyclic operating characteristics because of the ash removal, and combustion cycles that are erratic and difficult to control. Because of the slow response time of the system and the cyclic nature of operation, pile burners are not considered for load-following operations. The advantage of the pile burner is its simplicity and ability to handle wet, dirty fuels.

Stoker combustors (Hollenbacher 1992), improve on operation of the pile burners by providing a moving grate which permits continuous ash collection, thus eliminating the cyclic operation characteristic of traditional pile burners. In addition, the fuel is spread more evenly, normally by a pneumatic stoker, and in a thinner layer in the combustion zone, giving more efficient combustion. Stoker fired boilers were first introduced in the 1920's for coal; and in the late 1940's the Detroit Stoker Company installed the first traveling grate spreader stoker boiler for wood. In the basic stoker design the bottom of the furnace is a moving grate which is cooled by underfire air. Underfire air rate defines the maximum temperature of the grate and thus the allowable moisture content of the feed. More modern designs include the Kabliz grate, a sloping reciprocating water-cooled grate. Reciprocating grates are attractive because of simplicity and low fly ash carryover. Combustion is completed by the use of overfire air. Furnace wall configurations include straight and bull nose water walls. Vendors include Zurn, Foster Wheeler, and Babcock & Wilcox.

In a gas-solid fluidized bed, a stream of gas passes upwards through a bed of free-flowing granular materials in which the gas velocity is large enough that the solid particles are widely separated and circulate freely throughout the bed. During overall circulation of the bed, transient streams of gas flow upwards in channels containing few solids, and clumps or masses of solids flow downwards (Perry and Chilton 1973). The fluidized bed looks like a boiling liquid and has the physical properties of a fluid. In fluidized-bed combustion of biomass, the gas is air and the bed is usually sand or limestone. The air acts as the fluidizing medium and is the oxidant for biomass combustion. A fluidized-bed combustor is a vessel with dimensions such that the superficial velocity of the gas maintains the bed in a fluidized condition at the bottom of the vessel. A change in cross-sectional area above the bed lowers the superficial gas velocity below fluidization velocity to maintain bed inventory and acts as a disengaging zone. Overfire air is normally introduced in the disengaging zone. To obtain the total desired gas-phase residence time for complete combustion and heat transfer to the boiler walls the larger cross-sectional area zone is extended and is usually referred to as the freeboard. A cyclone is used to either return fines to the bed or to removes ash-rich fines from the system. The bed is fluidized by a gas distribution manifold or series of sparge tubes (Hansen 1992).

If the air flow of a bubbling fluid bed is increased, the air bubbles become larger, forming large voids in the bed and entraining substantial amounts of solids. This type of bed is referred to as a turbulent fluid bed (Babcock and Wilcox 1992). In a circulating fluid bed the turbulent bed solids are collected, separated from the gas, and returned to the bed, forming a solids circulation loop. A circulating fluid bed can be differentiated from a bubbling fluid bed in that there is no distinct separation between the dense solids zone and the dilute solids zone. The residence time of the solids in a circulating fluid bed is determined by the solids circulation rate, the attritibility of the solids, and the collection efficiency of the solids separation device. As with bubbling fluid beds, the primary driving force for development of circulating fluid beds in the United States is emissions. The uniform, low combustion temperature gives low NO_x emissions. In a circulating fluid bed, with its need for introduction of solids to maintain bed inventory, it is easy to introduce a sorbent solid, such as limestone or dolomite, to control SO_2 emissions without the need for back-end sulfur removal equipment. Circulating fluid bed temperatures are maintained at about 870°C, which helps to optimize the limestone-sulfur reactions (Tampella Power 1992). The major manufacturers of circulating fluid bed boilers for biomass are Combustion Engineering (CE-Lurgi), B&W-Studsvik, Ahlstrom Pyropower (Foster Wheeler) and Gotaverken. A number of plants have been built in the 25 MW size range, primarily in California.

The suspension burning of pulverized wood in dedicated biomass boilers is a fairly recent development and is practiced in relatively few installations. Suspension burning has also been accomplished in lime kilns (MacCallum 1992) and is being investigated by the utility industry for cofiring applications (Tillman et al 1994). For successful suspension firing, a feed moisture content of less than 15% (Hollenbacher 1992) and a particle size less than 0.0015 m (MacCallum 1992) give higher boiler efficiencies, up to 80%, than firing wet wood chips, 50-55% moisture, in a stoker grate or fluid bed, at 65% efficiency. The higher efficiency also results in smaller furnace size. Offsetting the higher efficiency is the cost and power consumption of drying and comminution. In addition, special burners need to be used. Burners developed for suspension firing include scroll cyclonic burners and vertical-cylindrical burners (Hollenbacher 1992). Installations include the Oxford Energy, 27 MW facility at Williams, California (Hollenbacher 1992); the ASSI Lövholmen Linerboard Mill in Piteå, Finland (Westerberg 1981); the Klabin do Parana mill in Monte Alegre, Brazil (MacCallum 1992); and the E.B. Eddy Mill in Espanola Ontario (MacCallum 1992).

The base technology is a commercially available stoker-grate biomass plant constructed in the mid-1980's (EPRI 1993b), and is representative of modern biomass plants with an efficiency of about 23%. Plant efficiency of the stoker plant increased in the case study to 30% through the use of a dryer and steam cycle efficiency increases, e.g. higher pressure, higher temperature and reheat.

The feedstock used is assumed to be a mixture of pine and oak (40% pine - 60% oak) with 50% moisture content. This feed mixture was also used in the gasification analysis. For cofiring a mixture residues was assumed, with blending to reduce the moisture content to 21.5%, thus eliminating the need for a dryer. An analysis of the two feeds is given in Table 5.2.

The starting case is based on EPRI report TR-102107, v2 (Wiltsee and Hughes 1995), for the Burlington, VT, McNeil Station. Wood heating values are about 10 MJ/kg on a wet basis and 20 MJ/ kg on a dry basis; these values are about 40% and 80% of coal (24.78 MJ/kg [AEO97 1996]), respectively. The name plate efficiency of the McNeil station is 25%, while the Biopower model (EPRI 1995) gives 23.0%. An average of 24% was used as the starting point for the case study.

The RETC97 capital and operating costs were updated to 2001 dollars using the Marshall and Swift Index (Marshall and Swift ------), and plant costs were updated by adding a dryer (Craig and Mann 1996). Capital and operating costs for other plant sizes were scaled from the 50 MW values using a 0.7 scaling factor. Peters and Timmerhaus (Peters and Timmerhaus 1980) state, “It is often necessary to estimate the cost of a piece of equipment when no cost data are available for the particular size of operational

Table 5.2 Feedstock Composition for direct combustion and gasification

Component	Pine		Oak	
	5%M	50%M	5%M	50%M
C, wt%	50.45	26.55	47.65	25.08
H	5.74	3.02	5.72	3.01
N	0.16	0.09	0.09	0.05
O	37.34	19.66	41.17	21.65
S	0.02	0.01	0.01	0.01
Cl	0.03	0.01	0.01	0.01
Moisture	5.00	50.00	5.00	50.00
Ash	1.26	0.67	0.35	0.19
MJ/kg (wet)	19.72	10.38	18.92	9.96
MJ/kg (dry)	20.76	20.76	19.92	19.92

capacity involved. Good results can be obtained by using the logarithmic relationship known as the 'six-tenths-factor rule,' if the new piece of equipment is similar to one of another capacity for which cost data are available. According to this rule, if the cost of a given unit at one capacity is known, the cost of a similar unit with X times the capacity of the first is approximately $(X)^{0.6}$ times the cost of the initial unit." Valle-Riesta (Valle-Riesta 1983) states " A logical consequence of the 'sixth-tenths-factor' rule for characterizing the relationship between equipment capacity and cost is that a similar relationship should hold for the direct fixed capital of specific plants.....In point of fact, the capacity exponent for plants, on the average, turns out to be closer to 0.7." The exception to this rule happens when plant capacity is increased by change in efficiency, not change in equipment size. In this case, capital cost in dollars remains constant, and capital cost in $/kW decreases in proportion to efficiency increase.

The electrical substation is part of the general plant facilities, and is not separated out in the factor analysis. The convention follows that used in the EPRI Technical Assessment Guide (EPRI 1993a), as follows "It also includes the high-voltage bushing of the generation step-up transformer but not the switchyard and associated transmission lines. The transmission lines are generally influenced by transmission system-specific conditions and hence are not included in the cost estimate." A summary of capital and operating costs is given in Table 5.3.

To estimate plant performance as a CHP facility the steam conditions from the Biopower model (8.72 MPa and 510°C) were used in an ASPEN™ steam turbine simulation to estimate steam turbine performance in three modes of operation—as a condensing turbine (comparable to the RETC97 electricity case), as a backpressure turbine, and as an extraction turbine. The steam efficiency was assumed to be 80%. The extraction turbine case was used in CHP performance estimates. The use of the extraction turbine gave a heat (H) to power ratio (P) of 1.44, as shown in Table 5.12. A summary of turbine performance for a 50 MW_e equivalent facility is given in Table 5.4. To convert to net plant efficiency a parasitic load of 5 MW_e is subtracted from gross electricity production.

Table 5.3: Biomass Direct Combustion Plant Capital and Operating Costs (excluding feed)

Indicator Name	Units	M.F.	Scale Factor	Base	2001 Base	2001 25 MW	2001 75 MW	2001 100MW
				1996	2001	2001	2001	2001
M&S index				1039	1092			
Plant Size Net	MW			50	50	25	75	100
Gross	MW				55	28	83	110
General Performance Indicators								
Capacity Factor	%			80%	90%	90%	90%	90%
Efficiency	%			24.0%	30%	30%	30%	30%
Net Heat Rate	Btu/kWh			14,234	11,373	11,373	11,373	11,373
Annual Energy Delivery	GWh/yr			350	394	197	591	788
Capital Cost	$000							
Fuel Preparation			0.7	9,052	8,373	5,154	11,121	13,602
Dryer			0.7	-	4,418	2,719	5,868	7,177
Boiler			0.7	22,215	20,549	12,649	27,293	33,381
Baghouse and Cooling Tower			0.7	1,456	1,347	829	1,789	2,188
Boiler feedwater/ deaerator			0.7	2,784	2,575	1,585	3,420	4,183
Steam turbine/generator			0.7	7,407	6,851	4,218	9,100	11,130
Cooling water system			0.7	3,312	3,064	1,886	4,069	4,977
Balance of plant			0.7	13,641	12,618	7,767	16,759	20,498
Subtotal (A)				59,867	59,794	36,808	79,419	97,136
General Plant Facilities (B)			0.7	15,498	14,336	8,825	19,040	23,288
Engineering Fee, k*(A+B)		0.1		7,537	7,413	4,563	9,846	12,042
Process/project contingency		0.15		11,305	11,119	6,845	14,769	18,064
Total Plant Cost (TPC)				94,206	92,662	57,040	123,074	150,530
AFUDC				2,826	2,780	1,711	3,692	4,516
Total Plant Investment (TPI)				97,032	95,442	58,752	126,766	155,046
Prepaid Royalties			1	-				
Initial Cat & Chem Inventory			1	110	76	38	115	153
Startup costs			1	2,653	1,842	921	2764	3685
Inventory Capital			0.7	559	433	217	650	866
Land, 100 acres@ $7,250/acre			1	725	725	725	725	725
Total Capital Requirement (TCR)				**101,079**	**98,519**	**60,652**	**131,019**	**160,475**

Indicator Name	Units	M.F.	Scale Factor	Base	2001 Base	2001 25 MW	2001 75 MW	2001 100MW
Capital	$/kW							
Fuel Preparation				181	167	206	148	136
Dryer				-	88	109	78	72
Boiler				444	411	506	364	334
Baghouse and Cooling Tower				29	27	33	24	22
Boiler feedwater/ deaerator				56	52	63	46	42
Steam turbine/generator				148	137	169	121	111
Cooling water system				66	61	75	54	50
Balance of plant				273	252	311	223	205
Subtotal (A)				1,197	1,196	1,472	1,059	971
General Plant Facilities (B)				310	287	353	254	233
Engineering Fee, k*(A+B)				151	148	183	131	120
Process/project contingency				226	222	274	197	181
Total Plant Cost (TPC)				1,884	1,853	2,282	1,641	1,505
AFUDC				57	56	68	49	45
Total Plant Investment (TPI)				1,941	1,909	2,350	1,690	1,550
Prepaid Royalties				-	-	-	-	-
Initial Cat & Chem Inventory				2	2	2	2	2
Startup costs				53	37	37	37	37
Inventory Capital				11	9	9	9	9
Land, 100 acres@ $7,250/acre				15	15	29	10	7
Total Capital Requirement (TCR)	**$/kW**			**2,022**	**1,970**	**2,426**	**1,747**	**1,605**
Operating					1,628	814	2,442	3,256
Supervision and Clerical					408	408	408	408
Maintenance Labor and Material Costs @ 1.8% of TPC					1,668	834	2,502	3,336
Total Fixed Costs	K$/a				3,704	2,056	5,351	6,999
Variable Costs (without feed)								
Labor					1,349	675	2,024	2,699
Maintenance Labor and Material Costs					768	384	1,152	1,536
Total Variable Costs	K$/a				2,118	1,059	3,176	4,235
Consumables								
Chemicals					670	335	1,006	1,341
Water					169	85	254	338
Solids/ash disposal					182	91	273	364
Ammonia					106	53	160	213
Total Consumables	K$/a				1,128	564	1,692	2,256
Feed estimates								
Capacity Factor	%				90%	90%	90%	90%
Feed heating value	Mbtu/ton				17	17	17	17

Indicator Name	Units	M.F.	Scale Factor	Base	2001 Base	2001 25 MW	2001 75 MW	2001 100MW
Process efficiency	%				30%	30%	30%	30%
Feed req'd	Mbtu/MWh				11	11	11	11
Generation	MWh/a				394,200	197,100	591,300	788,400
Annual feed	Mbtu/a				4,483,368	2,241,684	6,725,052	8,966,736
	ton/a				263,728	131,864	395,591	527,455
Total Operating Costs	**K$/a**				**6,949**	**3,678**	**10,220**	**13,490**
Annual Operating Costs								
Fixed Costs								
Operating					0.0041	0.0041	0.0041	0.0041
Supervision and Clerical					0.0010	0.0021	0.0007	0.0005
Maintenance Labor and Material Costs @ 1.8% of TPC					0.0042	0.0042	0.0042	0.0042
Total Fixed Costs	$/kWh				0.0094	0.0104	0.0091	0.0089
Variable Costs (without feed)								
Labor					0.0034	0.0034	0.0034	0.0034
Maintenance Labor and Material Costs					0.0019	0.0019	0.0019	0.0019
Total Variable Costs	$/kWh				0.0054	0.0054	0.0054	0.0054
Consumables								
Chemicals					0.0017	0.0017	0.0017	0.0017
Water					0.0004	0.0004	0.0004	0.0004
Solids/ash disposal					0.0005	0.0005	0.0005	0.0005
Ammonia					0.0003	0.0003	0.0003	0.0003
Total Consumables	$/kWh				0.0029	0.0029	0.0029	0.0029
Total Operating Costs	**$/kWh**				**0.0176**	**0.0186**	**0.0173**	**0.0172**

Table 5.4: Direct-fired Combustion Cogeneration Gross Plant Efficiencies

	Temp (°C)	Temp (°F)	Pressure (MPa)	Pressure (psia)	Flow (kg/s)	Flow (lb/hr)	Quality (%)	Electricity (MW_e)	Steam (MW_t)	HP Steam (MW_t)	Efficiency Electric (%)	Efficiency Thermal (%)	Efficiency Steam (%)
Condensing Turbine													
Inlet	510	848	8.720	1,279	54.00	428,610	100	55	0	172.5	31.9	0	31.9
Extraction[a]													
Outlet	54	130	0.0152	2.2	54.00	428,610	91						
Backpressure Turbine													
Inlet	510	848	8.720	1,279	54.00	428,610	100	24.2	137.5	172.5	14.0	93.7	79.2
Extraction[a]													
Outlet	266	510	1.140	150	54.00	428,610	91						
Extraction Turbine													
Inlet	510	848	8.720	1,279	54.00	428,610	100	39.6	69.2	172.5	22.9	40.1	63.0
Extraction[a]	266	510	1.140	150	27.00	214,305	100						
Outlet	54	130	0.0152	2.2	27.00	214,305	91						

[a] Doesn't include process use extraction

Gasification

This discussion characterizes a biomass-based power plant that utilizes a gasification combined cycle (GCC) system as depicted in Figure 5.2. Generally speaking, the conversion of biomass to a low- or medium-heating-value gaseous fuel (biomass gasification) involves two processes. The first process, pyrolysis, releases the volatile components of the fuel at temperatures below 600°C (1112°F) via a set of complex reactions. Included in these volatile vapors are hydrocarbon gases, hydrogen, carbon monoxide, carbon dioxide, tars, and water vapor. Because biomass fuels tend to have more volatile components (70 - 86% on a dry basis) than coal (30%), pyrolysis plays a proportionally larger role in biomass gasification than in coal gasification. The by-products of pyrolysis that are not vaporized are referred to as char and consist mainly of fixed carbon and ash. In the second gasification process, char conversion, the carbon remaining after pyrolysis undergoes the classic gasification reaction (i.e. steam + carbon) and/or combustion (carbon + oxygen). It is this latter, combustion, reaction that provides the heat energy required to drive the pyrolysis and char gasification reactions. Due to its high reactivity (as compared to coal and other solid fuels), all of the biomass feed, including char, is normally converted in a single pass through a gasifier system.

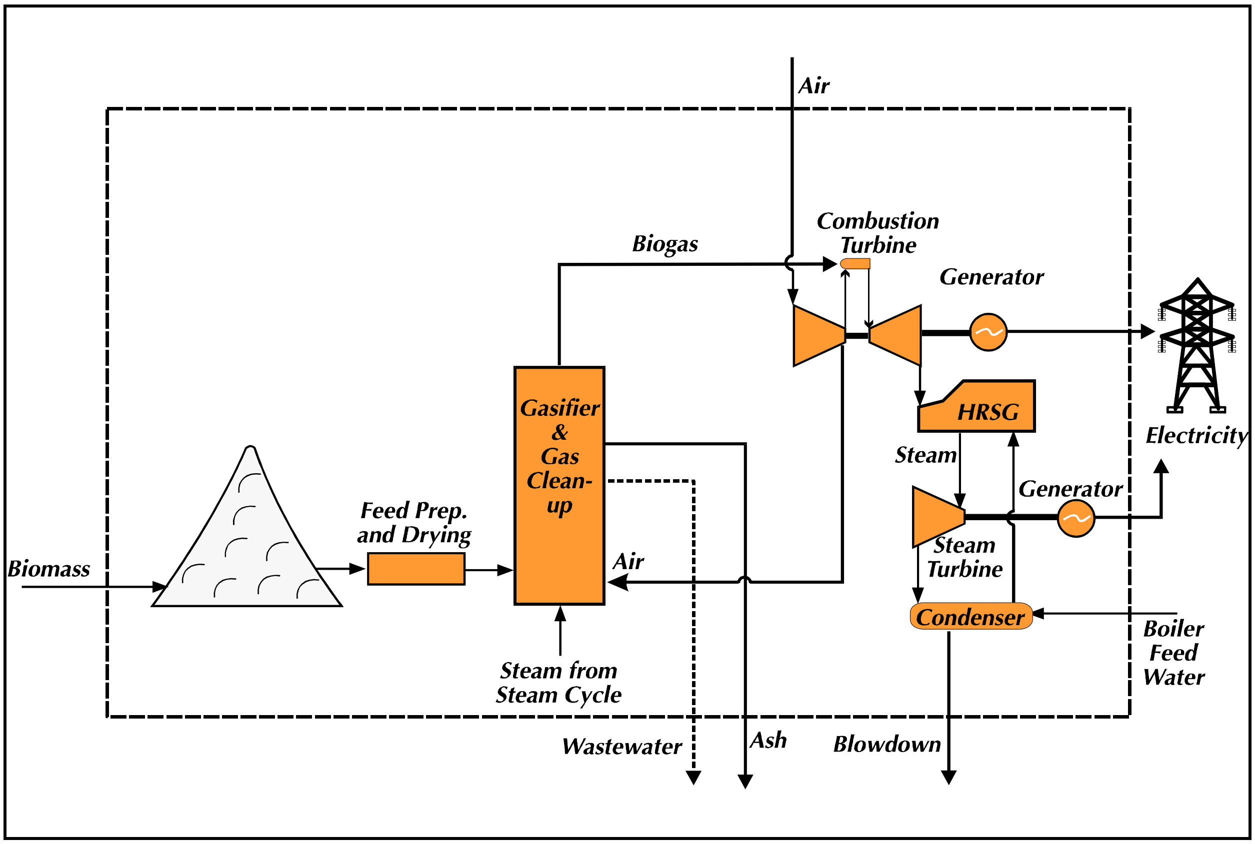


Figure 5.2: Generic Biomass Gasification Combined Cycle System

Depending on the type of gasifier used, the above reactions can take place in the same reactor vessel or separate vessels. These gasifier types are typically referred to as direct (pyrolysis, gasification, and combustion take place in one vessel) and indirect (pyrolysis and gasification in one vessel, combustion in a separate vessel). In direct gasification, air and sometimes steam are directly introduced to the single gasifier vessel (Figures 5.3 and 5.5). In indirect gasification, an inert heat transfer medium such as sand carries heat generated in the combustor to the gasifier to drive the pyrolysis and char gasification

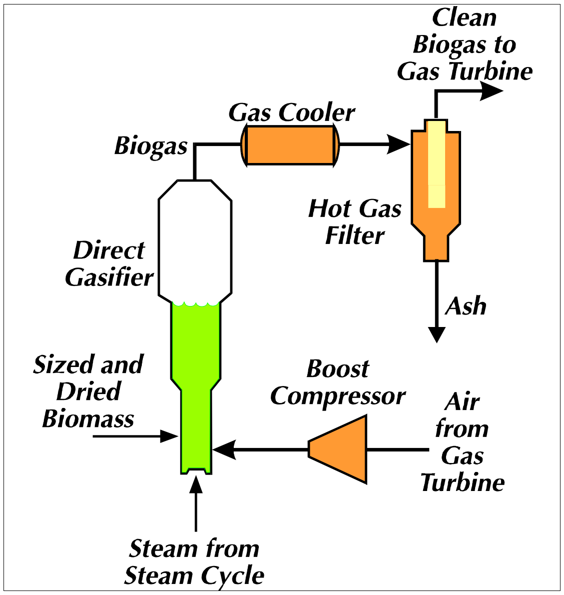


Figure 5.3: High Pressure Direct Gasifier

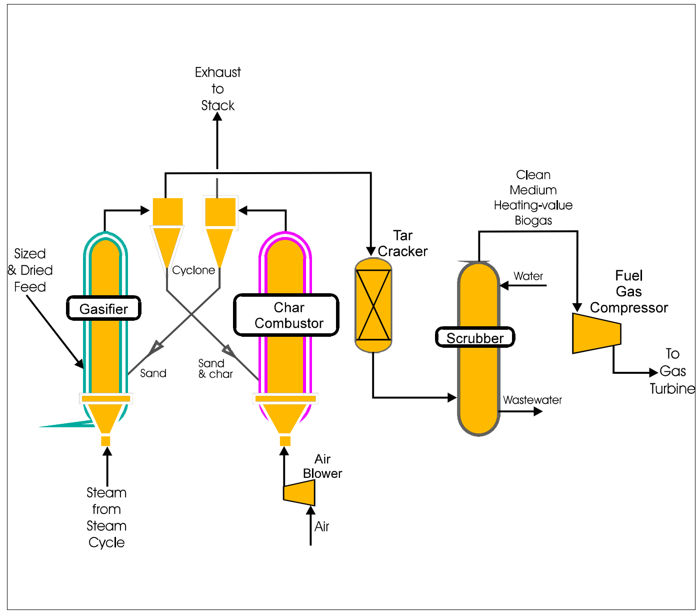


Figure 5.4: Indirect Gasifier

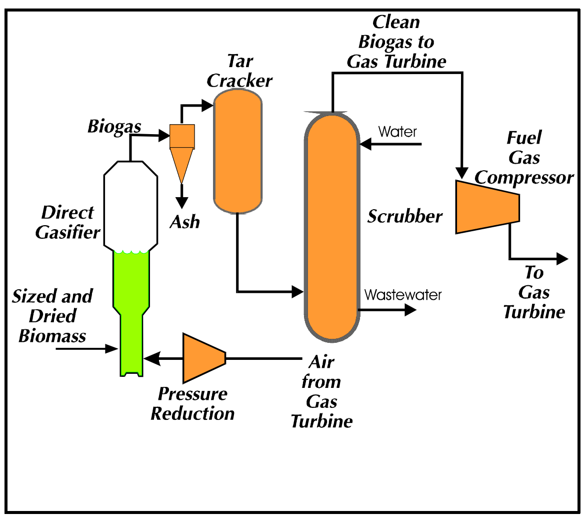


Figure 5.5: Low Pressure Direct Gasifier

reactions (Figure 5.4). Currently, indirect gasification systems operate near atmospheric pressure. Direct gasification systems have been demonstrated at both elevated (Figure 5.4) and atmospheric pressures (Figure 5.3). Any one of the gasifier systems shown in Figures 5.3 - 5.5 can be utilized in the generic gasifier block represented in the main system diagram above and have been utilized in a least one recent system design study (NSP 1995; Weyerhaeuser 1995; Craig and Mann 1996; EPRI 1995).

There are several practical implications of each gasifier type. Because of the diluent effect of nitrogen in air, fuel gas from a direct gasifier is of low heating value (5.6 - 7.5 MJ/Nm^3). This low heat content in turn requires an increased fuel flow to the gas turbine. Consequently, to maintain the total (fuel + air) mass flow through the turbine within design limits, an air bleed is usually taken from the gas turbine compressor and used in the gasifier. This bleed air is either boosted slightly in pressure or expanded to near atmospheric pressure depending on the operating pressure of the direct gasifier.

Because the fuel-producing reactions in an indirect gasifier take place in a separate vessel, the resulting fuel gas is free of nitrogen diluent and is of medium heating value (13 - 18.7 MJ/Nm^3). This heat content is sufficiently close to that of natural gas (approx. 38 MJ/Nm^3) that fuel gas from an indirect gasifier can be used in an unmodified gas turbine without air bleed.

Gasifier operating pressure impacts not only equipment cost and size but the interfaces to the rest of the power plant including the necessary cleanup systems. Since gas turbines operate at elevated pressures, the fuel gas generated by low pressure gasifiers must be compressed. This favors low temperature gas cleaning since the fuel gas must be cooled prior to compression in any case. Air for a low pressure gasifier can be extracted from the gas turbine and reduced in pressure (direct, low pressure gasifier) or supplied independently (indirect gasifier). High pressure gasification favors hot, pressurized cleanup of the fuel gas and supply to the gas turbine combustor at high temperature (circa 538ºC or 1000ºF) and sufficiently high pressure for flow control and combustor pressure drop. Air for a high pressure, direct gasifier is extracted from the gas turbine and boosted in pressure prior to introduction to the gasifier.

Cooling, cold cleanup, and fuel gas compression add equipment to an indirect gasifier system and reduce its efficiency by up to 10% (Craig and Mann 1996, Marrison and Larson 1995). Gasifier and gas cleanup vessels rated for high pressure operation and more elaborate feed systems, however, add cost and complexity to high pressure gasification systems despite their higher efficiency. Results from several recent studies (NSP 1995, Weyerhaeuser 1995, Craig and Mann 1996, Marrison and Larson 1995) indicate that, at the current, preliminary grade of estimates (as defined by EPRI TAG, 1993) being performed, there is little discernable difference in cost of electricity (COE) between systems employing high and low pressure gasification.

For the purposes of this analysis, a high-pressure, direct gasification system as shown in Figure 5.3 was selected. The resulting system is very similar to that evaluated in a pre-feasibility study conducted by Northern States Power under subcontract AAE-5-14456-01 for NREL and EPRI, reported in NREL/TP-430-20517 (NSP 1995). This study examined a 75 MW$_e$ power plant that would gasify alfalfa stems to provide electricity to the Northern States Power Company and sell the leaf co-product for animal feed. A departure from the NSP study is the use here of wood as the biomass feedstock. Wood feedstock allows for a more generic plant representation. Alfalfa separation and leaf meal processing steps in the original NSP study would have added complexity and cost to the plant and have complicated the economic analysis.

Following receipt of wood chips at the plant, they are screened and hogged to a proper size consistency, and dried in a rotary drum dryer. Dried wood is conveyed to storage silos adjacent to the gasifier building. It is then weighed and transferred to a lockhopper/screw feeder system and is fed into the fluidized bed gasifier. The gasifier vendor selected for the NSP study was Tampella Power Systems (now Carbona) who have developed a commercial version of the IGT RENUGAS™ gasifier. A dolomite feed system is also provided to maintain the inventory of inert material in the bed. In the gasifier, the biomass is gasified at temperatures between 843ºC (1550ºF) and 954ºC (1750ºF). The fluidizing and gasifying medium is a mixture of air and steam. As shown above, air is extracted from the compressor section of the gas turbine and fed into the gasifier through a boost compressor. Gasification steam is extracted from the steam cycle. The gasifier operates as a so-called spouted bed with intensive circulation of solids from top to bottom, which guarantees rapid gasification and maximizes tar cracking.

Fuel gases exiting the gasifier are cooled in the product gas cooler to approximately 538ºC (1000ºF). In addition to protecting the fuel flow control valve, this cooling causes the vapor-phase alkali species present in the fuel gas, which could damage the gas turbine, to condense, congeal, and deposit on the fine particulate matter carried over from the gasifier. The combined particulate matter and alkali species are next removed in a Westinghouse hot ceramic candle filter unit to levels within gas turbine tolerances.

Since biomass in general and wood in particular is very low in sulfur, a sulfur removal step is not necessary prior to combustion in the gas turbine. Hot cleanup of the fuel gas also minimizes wastewater generation from this step of gas processing.

The fuel gas is combusted in a Westinghouse "ECONOPAC" 251B12 gas turbine producing electric power and a high temperature exhaust stream. A heat recovery steam generator (HRSG) is employed to recover this heat to generate high temperature, high pressure steam that is then expanded in a steam turbine to produce additional power. Steam for the gasifier is extracted from the steam cycle. As noted above, the total net electricity output from this system is 75 MW_e. The following cost and performance estimates, Table 5.5, were scaled to 150 MW using the 0.7 rule. It is worth noting that rapid developments are also being made in smaller turbine sizes as well, and the industrial and cogeneration markets (10 - 50 MW_e output) should not be ignored.

As mentioned earlier, several gasifier configurations could have been considered. Converting solid biomass into a gaseous fuel with suitable heating value creates the opportunity to integrate biomass gasifiers with the gas turbine cycles such as the combined gas and steam cycle depicted above. Close coupling of gasification and the power system increases overall conversion efficiency by utilizing both the thermal and chemical energy of hot product gases to fuel the power cycle. Combined cycles, with their high efficiency and low emission characteristics, are a prime choice for biomass gasification systems.

To estimate plant performance as a CHP facility the steam conditions given for year 2000 technology were used in an ASPEN™ steam turbine simulation to estimate steam turbine performance in three modes of operation—as a condensing turbine (comparable to the RETC97 electricity case), as a back-pressure turbine, and as an extraction turbine. The steam efficiency was assumed to be 80%. The back-pressure turbine case was used in CHP performance estimates. The back-pressure turbine was chosen to give a H/P approximately the same as the direct combustion case. The gasification H/P was 1.60 compared to 1.44 for direct combustion. A summary of turbine performance for a 75 MW_e equivalent facility is given in Table 5.6. To convert to net plant efficiency a parasitic load of 6.7 MW_e was subtracted from gross electricity production, and the gas turbine production of 51 MW_e was added.

Table 5.5: Biomass Gasification Capital and Operating Costs, Excluding Feed

Indicator Name	Units	M.F.	S.F.				
				1996	1996	2001	2001
M&S index				1039.1	1039.1	1092	1092
Plant Size Net	MW			75	75	75	150
Gross	MW						
General Performance Indicators							
Capacity Factor	%			80%	90%	90%	90%
Efficiency	%			36%	36%	36%	36%
Net Heat Rate	Btu/kWh			9,483	9,483	9,483	9,483
Annual Energy Delivery	GWh/yr			526	591	591	1,183
Capital Cost	$000						
Fuel Preparation			0.7	7,575	7,575	7,961	12,932
Gasifier			0.7	25,950	25,950	27,271	44,302
Gas Turbine			0.7	14,850	14,850	15,606	25,352
Steam Turbine			0.7	3,300	3,300	3,468	5,634
Balance of Plant			0.7	11,025	11,025	11,586	18,822
Control System			0.7	600	600	631	1,024
Hot gas Cleanup			0.7	2,325	2,325	2,443	3,969
Installation			0.7	9,900	9,900	10,404	16,901
Turbine Building			0.7	450	450	473	768
Waste Pond			1	150	150	158	315
Subtotal (A)				76,125	76,125	80,000	130,020
General Plant Facilities (B)		0.1	0.7	7,613	7,613	8,000	13,002
Engineering Fee, k*(A+B)		0.1		8,374	8,374	8,800	14,302
Process/project contingency		0.15		12,561	12,561	13,200	21,453
Total Plant Cost (TPC)				104,672	104,672	110,001	178,778
AFUDC					-	-	
Total Plant Investment (TPI)				104,672	104,672	110,001	178,778
Prepaid Royalties			1				
Initial Cat & Chem Inventory			1				
Startup costs			1	4,200	4,200	4,414	8,828
Inventory Capital			0.7	750	750	788	1,576
Land, 100 acres@ $7,250/acre				750	750	788	1,576
				5700	5,700	5,990	5990

Indicator Name	Units	M.F.	S.F.				
Total Capital Requirement (TCR)				**110,372**	110,372	115,991	**196,749**
Capital	$/kW						
Fuel Preparation				101	101	106	86
Gasifier				346	346	364	295
Gas Turbine				198	198	208	169
Steam Turbine				44	44	46	38
Balance of Plant				147	147	154	125
Control System				8	8	8	7
Hot gas Cleanup				31	31	33	26
Installation				132	132	139	113
Turbine Building				6	6	6	5
Waste Pond				2	2	2	2
Subtotal (A)				1,015	1,015	1,067	867
General Plant Facilities (B)		0.1		102	102	107	87
Engineering Fee, k*(A+B)		0.1		112	112	117	95
Process/project contingency		0.15		167	167	176	143
Total Plant Cost (TPC)				1,396	1,396	1,467	1,192
AFUDC						-	-
Total Plant Investment (TPI)				1,396	1,396	1,467	1,192
Prepaid Royalties							
Initial Cat & Chem Inventory							
Startup costs				56	56	59	59
Inventory Capital				10	10	11	11
Land, 100 acres@ $7,250/acre				10	10	11	11
						80	40
Total Capital Requirement (TCR)	**$/kW**			**1,472**	**1,472**	**1,626**	**1,312**
Annual Operating Costs							
Fixed Costs							
Operating				535	535	562	1,124
Supervision and Clerical				435	435	457	457
Maintenance Labor and Material Costs			0.7	2,285	2,285	2,402	3,901
Total Fixed Costs	K$/a			3,255	3,255	3,421	5,482
Variable Costs (without feed)							
Labor				1,787	2,010	2,113	4,021
Maintenance Labor and Material Costs				315	355	373	710
Total Variable Costs	K$/a			2,102	2,365	2,486	4,730

Indicator Name	Units	M.F.	S.F.				
Consumables							
Chemicals				210	237	249	473
Water				315	355	373	710
Solids/ash disposal				158	177	186	355
Ammonia							
Total Consumables	K$/a			683	769	808	1,537
Total Operating Costs	**K$/a**			**6,041**	**6,389**	**6,714**	**11,750**
Annual Operating Costs							
Fixed Costs							
Operating				0.0010	0.0009	0.0010	0.00095
Supervision and Clerical				0.0008	0.0007	0.0008	0.00039
Maintenance Labor and Material Costs @ 1.8% of TPC				0.0043	0.0039	0.0041	0.00330
Total Fixed Costs	$/kWh			0.0062	0.0055	0.0058	0.00464
Variable Costs (without feed)							
Labor				0.0034	0.0034	0.0036	0.0034
Maintenance Labor and Material Costs				0.0006	0.0006	0.0006	0.0006
Total Variable Costs	$/kWh			0.004	0.004	0.0042	0.004
Consumables							
Chemicals				0.0004	0.0004	0.0004	0.0004
Water				0.0006	0.0006	0.0006	0.0006
Solids/ash disposal				0.0003	0.0003	0.0003	0.0003
Ammonia							
Total Consumables	$/kWh			0.0013	0.0013	0.0014	0.0013
Total Operating Costs	**$/kWh**			0.0115	0.0108	0.0114	0.0099

Table 5.6: Gasification Cogeneration Gross Steam Turbine Efficiencies

	Temp		Pressure		Flow		Quality	Electricity	Steam	Efficiency		
	(°C)	(°F)	(MPa)	(psia)	(kg/s)	(lb/hr)	(%)	(MW_e)	(MW_t)	Electric (%)	Thermal (%)	Steam (%)
Condensing Turbine												
Inlet	468	874	5.860	865	32.64	259,045	100	31	0	30.4	0	30.4
Extraction[a]												
Outlet	54	130	0.0152	2.2	32.64	259,045	92					
Backpressure Turbine												
Inlet	468	874	5.860	865	32.64	259,045	100	12	84	11.8	82.4	94.1
Extraction[a]												
Outlet	266	510	1.140	165	32.64	259,045	92					
Extraction Turbine												
Inlet	468	874	5.860	865	32.64	259,045	100	21	42	20.6	41.2	61.8
Extraction[a]	266	527	1.140	165	16.32	129,522	100					
Outlet	54	130	0.0152	2.2	16.32	129,522	92					

[a] Doesn't include process use extraction

Cofiring

Cofiring is the co-combustion of multiple fuels in the same boiler. Many coal- and oil-fired boilers at power stations have been retrofitted to permit multi-fuel flexibility. Biomass is well-suited for cofiring with other solid fuels, primarily coal, as an acid rain and greenhouse gas emission control strategy. Cofiring is a fuel-substitution option for existing fuel capacity, and is not a capacity expansion option. Cofiring utilizing biomass has been successfully demonstrated and is currently practiced in the full range of coal boiler types, including pulverized coal boilers, stokers, cyclones, and bubbling and circulating fluidized beds (Winslow et al. 1993). The system described here is for pulverized coal-fired boilers, which represent the majority of the current fleet of utility boilers in the U.S.; however, there are also significant opportunities for cofiring with biomass in stokers, cyclones, and fluidized bed boilers. Cofiring in an existing pulverized coal (PC) boiler will generally require minor modifications or additions to fuel handling, storage, and feed systems. An automated system capable of processing and storing sufficient biomass fuel in one shift for 24-hour use is needed to allow continuous cofiring. Typical biomass fuel receiving equipment will include truck scales and hydraulic tippers; however, tippers are not required if deliveries are made with self-unloading vans. New automated reclaiming equipment may be added or existing front-end loaders may be detailed for use to manage and reclaim biomass fuel. Conveyors will be added to transport fuel to the processing facility, with magnetic separators to remove spikes, nails, and tramp metal from the feedstock. Since biomass is

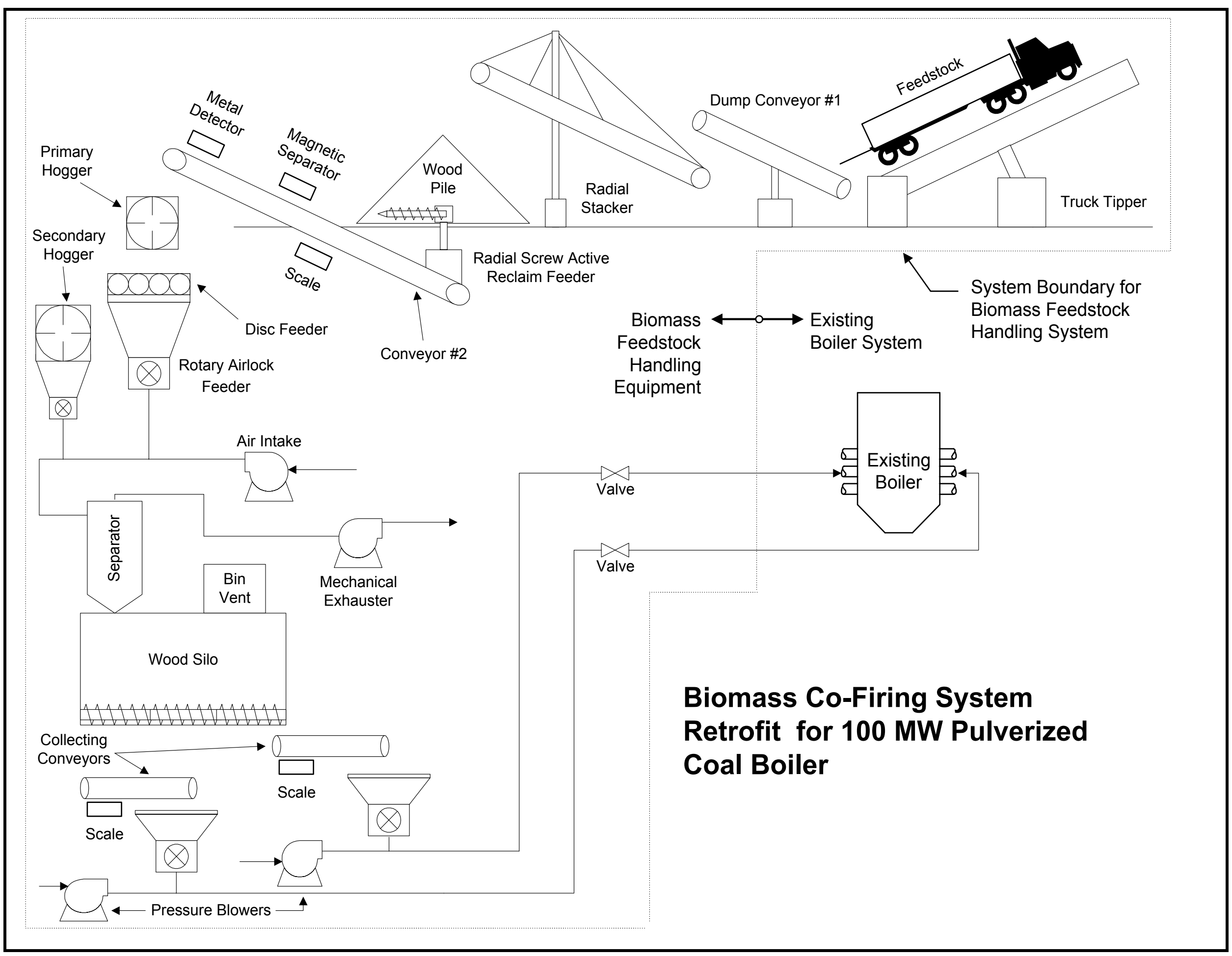


Figure 5.6: Biomass Co-firing System Retrofit for a Pulverized Coal Boiler

the “flexible” fuel at these facilities, a 5-day stockpile should be sufficient and avoids problems with longer term storage of biomass (Winslow et al. 1993).

Fuel processing requirements are dictated by the expected fuel sources, with incoming feedstocks varying from green whole chips up to 5 cm (2 inches) in size (or even larger tree trimmings) to fine dry sawdust requiring no additional processing. In addition to woody residues and crops, biomass fuel sources could include alfalfa stems, switchgrass, rice hulls, rice straws, stone fruit pits, and other materials (Hughes and Tillman 1996). For suspension firing in pulverized coal boilers, biomass fuel feedstocks should be reduced to a particle size of 6.4 mm (1/4 inch) or less with moisture levels under 25% MCW (Moisture Content, Wet basis) when firing in the range of 5% to 15% biomass on a heat input basis (Antares and Parsons 1996, Ebasco 1993). Equipment such as hoggers, hammer mills, spike rolls, and disc screens are required to properly size the feedstock. Other boiler types, such as stokers and fluidized beds are better suited to handle larger fuel particles. There must also be a biomass buffer storage and a fuel feed and metering system. Biomass is pneumatically conveyed from the storage silo and introduced into the boiler through existing injection ports. Introducing the biomass at the lowest level of burners helps to ensure complete burnout due to the scavenging effect of the upper level burners and the increased residence time in the boiler.

The system described here and shown in Figure 5.6 is designed for high percentage cofiring (>2% on a heat input basis) and, for that reason, requires a separate feed system for biomass which acts in parallel to the coal feed systems. Existing coal injection ports are modified to allow dedicated biomass injection during the cofiring mode of operation. For low percentage cofiring (<2% on a heat input basis), it may be possible to use existing coal pulverizers to process the biomass. If using existing pulverizers, the biomass is processed and conveyed to the boiler with the coal supply and is introduced into the boiler through the same injection ports as the coal. Using existing pulverizers could reduce capital costs by allowing the avoided purchase of dedicated biomass processing and handling equipment, but the level of cofiring on a percentage basis will be limited by pulverizer performance, biomass type, and excess pulverizer capacity. The suitability of existing pulverizers to process biomass with coal will vary depending on pulverizer type and biomass type. Atritta mills, for example, have significant capability to process biomass fuels (Hughes and Tillman 1996).

Drying equipment has been evaluated by many designers and recommended by some. Dryers are not included here for three reasons: (1) the benefit-to-cost ratio is almost always low, (2) the industrial fuel sources that supply most cofiring operations provide a moderately dry fuel (between 28% and 6% MCW), and (3) biomass is only a modest percentage of the fuel fired. Although drying equipment is not expected to be included initially, future designs may incorporate cost effective drying techniques (using boiler waste heat) to maintain plant efficiency while firing a broader range of feedstocks with higher moisture contents.

The current fleet of low cost, coal-fired, base load electricity generators are producing over 50% of the nation’s power supply (EIA 1996). With the Clean Air Act Amendments (CAAA) requiring reductions in emissions of acid rain precursors such as sulfur dioxide (SO_2) and nitrogen oxides (NO_x) from utility power plants, cofiring biomass at existing coal-fired power plants is viewed as one of many possible compliance options. In addition, cofiring using biomass fuels from sustainably grown dedicated energy crops is viewed as a possible option for reducing net emissions of carbon dioxide (CO_2), a greenhouse gas that contributes to global warming. Coupled with the need of the industrial sector to dispose of biomass residues (generally clean wood byproducts or remnants), biomass cofiring offers the potential for solving multiple problems at potentially modest investment costs. These opportunities have caught the interest of power companies in recent years.

Unlike coal, most forms of biomass contain very small amounts of sulfur. Hence, substitution of biomass for coal can result in significant reductions in sulfur dioxide emissions. Cofiring biomass with coal can allow power producers to earn sulfur dioxide (SO_2) emission allowances under section 404(f) of the

CAAA (U.S. House of Representatives 1990) (1 allowance = 1 ton of reduced SO_2 emissions = 0.91 metric ton of reduced SO_2 emissions). An allowance is earned for each ton of SO_2 emissions reduced. This section of the CAAA includes provisions for earning credits from SO_2 emissions avoided through energy conservation measures (i.e., demand side management or DSM) and renewable energy. In addition to any allowances which the producer saved by not emitting SO_2, two allowances can be given to the utility from an allowance reserve for every gigawatt-hour (GWh=10^6 kWh) produced by biomass in a co-fired boiler. These allowances may then be sold or traded to others who need them to remain in compliance with the CAAA.

Potential negative impacts associated with cofiring biomass fuels include: 1) the possibility for increased slagging and fouling on boiler surfaces when firing high alkali herbaceous biomass fuels such as switchgrass, and 2) the potential for reduced fly ash marketability due to concerns that commingled biomass and coal ash will not meet existing ASTM fly ash standards for concrete admixtures, a valuable fly ash market. These two issues are the subject of on-going research and investigation efforts. Two factors indicate that biomass cofiring (using sources of biomass such energy crops or residues from untreated wood) will have a negligible effect on the physical properties of coal fly ash. First, the mass of biomass relative to coal is small for cofiring applications, since biomass provides 15% or less of the heat input in the boiler. Second, combustion of most forms of biomass results in only half as much ash when compared to coal. Despite these factors, significant efforts will be required to ensure that commingled biomass and coal ash will be accepted by ASTM standards for concrete admixture applications.

Biomass cofiring is a retrofit application, primarily for coal-fired power plants. Biomass cofiring is applicable to most coal fired boilers used for power generation. A partial list of existing or planned utility applications is shown in Table 5.7. Recent DOE feasibility/demonstration projects are given in Table 5.8. Retrofits for coal-fired stokers, cyclones, and fluidized bed boilers are potentially simpler and less expensive than for pulverized coal. However, pulverized coal boilers are the most widely used steam generating system for coal-fired power generation in the U.S., and they represent most of plants affected by 1990 Clean Air Act Amendments (CAAA) provisions for reducing the emissions of SO_2 and NO_x from electric generating units.

The power plants characterized are pulverized coal plants which co-fire 15% biomass on a heat input basis. System capital and operating costs are assumed to be representative of plants which receive biomass via self-unloading vans and can utilize existing front-end loaders for receiving and pile management. The facilities are assumed to be located in a region where medium- to high-sulfur coal (0.8% by weight and greater) is used as a utility boiler fuel and where biomass residues are available for relatively low costs ($0.47/GJ, or $0.50/Mbtu). Areas with these characteristics include portions of the Northeast, Southeast, mid-Atlantic, and Midwest regions.

For each case, the performance of two systems is estimated. One is a pulverized coal power plant using only coal as a fuel source. These cases represent the plant operation prior to a biomass cofiring retrofit. The other case shows the performance of the same power plant operating with biomass cofiring. The tools used for this analysis were based on EPRI's BIOPOWER cofiring model (EPRI 1995). Input requirements for the model include ultimate analyses of the fuels (chemical composition of the fuels), capacity factor for the power plant, net station capacity, gross turbine heat rate, and percent excess air at which the plant operates. The technical input information used for the model were based on data from a representative Northeast power plant that intends to implement biomass cofiring. For a given biomass

Table 5.7: Existing or Planned Biomass Cofiring Applications (Winslow et al. 1996)

PLANT	FUEL	SIZE	TECHNOLOGY
Northern States Power Allen S. King Station Minneapolis, Minnesota	Coal/wood residues (lumber)	560 MWe	Cyclone
Otter Tail Power Co. Big Stone City, South Dakota	Coal/RDF/tires/ waste oil/ag. refuse	440 MWe	Cyclone
Tennessee Valley Authority Allen (1) & Paradise (2) Stations Memphis & Dunmore, Tennessee	1) Coal/wood residues and coal/wood/tires 2) Coal/wood residues	1) 176 MWe 2) 700 MWe	1) Cyclone 2) Cyclone
Elsam Grenaa Co-Generation Plant Grenaa, Denmark	Coal/straw	150 MWe	Circulating Fluidized Bed
Tacoma City & Light Tacoma Two Station Tacoma, Washington	Coal/RDF/wood residues	2 x 25 MWe	Bubbling Fluidized Bed
GPU GENCO Shawville Station Johnstown, Pennsylvania	Coal/wood residues	130 MWe	Pulverized Coal
IES Utilities Inc. Sixth Street (1) & Ottumwa (2) Stations Marshalltown, Iowa	1) Coal/agricultural residues 2) Coal/switchgrass	1) 3 Units 6-15 MWe 2) 714 MWe	1) Pulverized Coal 2) Pulverized Coal
Madison Gas & Electric Blount Street Station Madison, Wisconsin	Coal/switchgrass	50 MWe	Pulverized Coal
New York State Elec & Gas Greenidge Station Dresden, New York	Coal/wood residues and coal/energy crops (willow)	108 MWe	Pulverized Coal
Niagara Mohawk Power Corp. Dunkirk Station Dunkirk, New York	Coal/wood residues and coal/energy crops (willow)	91 MWe	Pulverized Coal
Tennessee Valley Authority Kingston Station Oakridge, Tennessee	Coal/wood residues	150 MWe	Pulverized Coal
EPON Centrale Gelderland Netherlands	Coal/wood residues (demolition)	602 MWe	Pulverized Coal
I/S Midtkraft Energy Co. Studstrupvaeket, Denmark	Coal/straw	150 MWe	Pulverized Coal
Uppsala Energi AB Uppsala, Sweden	Coal (peat)/ wood chips	200 MWe & 320 MWt	Pulverized Coal
New York State Elec & Gas Hickling (1) & Jennison (2) Stations Big Flats & Bainbridge, New York	Coal/wood residues and coal/tyres	1) 37.5 MWe 2) 37.5 MWe	1) Stoker 2) Stoker
Northern States Power Bay Front Station Ashland, Wisconsin	Coal/wood residues (forest)	2 x 17 MWe	Stoker

Table 5.8: Ongoing DOE Cofiring Feasibility/Demonstration Projects

Title	Organization	Description
Blending Biomass with Tire-Derived Fuel for Firing at Willow Island Generating Station	Allegeny Energy Supply Company	Researchers are demonstrating the blending of fuels for cofiring at the Willow Island Generating Station in West Virginia. Biomass fuels are expected to reduce harmful emissions form the power generating station.
Development of a Validated Model for Use in Minimizing NO_x Emissions and Maximizing Carbon Utilization When Cofiring Biomass With Coal	Southern Research Inst.	This project involves developing a computer model to calculate optimal energy and environmental benefits derived from cofiring biomass and coal.
Urban Wood/Coal Cofiring in the NIOSH Boiler Plant	University of Pittsburgh	The University of Pittsburgh is conducting cofiring demonstrations at the University's Bellefield boiler plant and at the NIOSH stoker boiler at the Bruceton Research Center.
Cofiring Biomass with Lignite Coal	Energy and Environmental Research Center, Grand Forks, ND	Tis demonstration wis cofiring wood waste with lignite coal at the North Dakota Penitentiary in Bismark.
Gasification-Based Biomass Cofiring Project	Northern Indiana Public Service Co.	The project is evaluating the feasibility of using wood waste, switchgrass, corn stover, non-recyclable paper and other related products to produce synthesis gas, and to fire the syngas in a generating that ordinarily fires natural gas.
Gasification-Based Cofiring Project	Nexant, LLC	Nexant will study the use of poultry litter in a biomass gasification cofiring demonstration at the Reid plant in Henderson, KY. This project will determine the optimum size at which gasifiers can be integrated, while maintaining boiler operation.
Calla Energy Biomass Gasification Cofiring Project		This project involves developing and Demonstrating as advanced version of the Gas Technology Institute RENUGAS™ biomass gasification technology to gasify biomass at a plant being built in Estill, KY. The gas will be used to produce steam and electricity in a 600-acre industrial park.
Feasibility Analysis for Installing a CFB Boiler for Cofiring Multiple Biofuels and Other Wastes with Coal	Pennsylvania State University	PSU is analyzing the installment of a state-of -the -art circulating fluidized bed boiler and ceramic emission control device, and is developing a test program to evaluate cofiring multiple biofuels and coal-based feedstocks.
Cofiring Coal: Feedlot and Litter Biomass Fuels in a Pulverized Fuel and Fixed-Bed Burners	Texas A&M University	Texas A&M University is investigating cattle feedlot and chicken litter biomass cofiring with coal to determine the optimum operating parameters and maximum combustion efficiency that can be achieved with the least emissions.
Cofiring Biomass at the University of North Dakota	University of North Dakota	This project is assessing local biomass resources available to the University and designing an economical feed system for the University's boiler.
Fuel-Lean Biomass Reburning in Coal-Fired Boilers	Iowa State University	ISU is examining the feasibility of adapting a commercially successful emissions reduction technology to herbaceous biomass when fired with coal.

cofiring rate, the model calculates thermal efficiency, change in net heat rate, coal and biomass consumption, and reduced SO_2 and CO_2 emissions.

The coal was assumed to contain 1.9% sulfur, compared to a 0.2% sulfur content for the biomass. Moisture contents were 7.2% for the coal and 21.5% for the biomass. The coal heating value was 31.75 MJ/kg (13,680 Btu/lb) (dry) and the biomass heating value was 19.10 MJ/kg (8,231 Btu/lb) (dry). These values for sulfur, moisture, and HHV were taken directly from tests conducted on the fuel supplies for the representative power plant. The resulting estimated net heat rate for coal-only operation is 10.93 MJ/kW (10,359 Btu/kW). This value is typical of high capacity factor coal boilers in the range from 100 MW to 400 MW, and was therefore assumed constant for all cases.

All system capital costs are due to the retrofit of an existing pulverized coal boiler to co-fire biomass. Costs shown in Table 5.9 are based on engineering specifications, including materials and sizing of major system components, from a feasibility study for a corresponding 10 MW (biomass power) biomass cofiring retrofit at an existing plant (Antares and Parsons 1996). The unit costs for the cofiring retrofit are expressed in $/kW of biomass power capacity, not total power capacity. Capital costs include costs for new equipment (e.g., fuel handling), boiler modifications, controls, engineering fees (10% of total process capital), civil / structural work including foundations and road ways, and a 15% contingency. Cost estimates for the example systems assume that front-end loaders and truck scales are already available at the plant for unloading and pile management. Costs also assume that live-bottom trucks are used for biomass delivery, allowing the avoidance of the purchase of a truck tipper. Land and substation (system interface) costs are zero because existing plant property and the existing substation will be utilized. Operation and maintenance costs, including fuel costs, are presented in the Table on an incremental basis. That is, each O&M cost component listed in the table represents the difference in that cost component when comparing biomass cofiring operation to coal-only operation. Negative costs, surrounded by parentheses in the table, represent a cost savings in the cofiring operation relative to coal-only operation. Updated plant performance indicators are given in Table 5.9, and updated capital and operating costs are given Table 5.10.

To estimate plant performance as a CHP facility, the biopower steam conditions (16.5 MPa and 538°C) were used in an ASPEN™ steam turbine simulation to estimate steam turbine performance in three modes of operation—as a condensing turbine (comparable to the RETC97 electricity case), as a backpressure turbine, and as an extraction turbine. The steam efficiency was assumed to be 80%. The extraction turbine case was used in CHP performance estimates. The use of the extraction turbine gave a heat (H) to power ratio (P) of 1.44, as shown in Table 5.12. A summary of turbine performance for a 100 MW_e equivalent facility is given in Table 5.11. To convert to net plant efficiency a parasitic load of 6 MW_e is subtracted from gross electricity production.

Table 5.9: Biomass Cofiring Performance Characteristics

Indicator Name	Units	Value	
Plant Size	MW	300	700
General Performance Indicators			
Capacity Factor	%	90	90
Total Electricity Generated	GWh/yr	2,365	5,518
Coal Moisture Content	%	7.2	7.2
Biomass Moisture Content	%	21.5	21.5
Coal-Only Performance Factors			
Plant Thermal Efficiency	%	32.9	32.9
Net Plant Heat Rate	KJ/kWh	10,929	10,929
	Btu/kWh	10,377	10,377
Net Power Capacity From Coal	MW	300	700
Annual Electricity From Coal	GWh/yr	2,365	5,518
Coal Consumption (wet)	tonnes/yr	877,550	2,047,617
Annual Heat Input From Coal	TJ/yr	25,847	60,310
Total Annual Heat Input	TJ/yr	25,847	60,310
Biomass Cofiring Performance Indicators			
Cofiring Rate (Heat Input From Biomass)	%	15	15
Plant Thermal Efficiency	%	32.5	32.5
Net Plant Heat Rate	kJ/kWh	11,066	25,821
	Btu/kWh	10,505	24,512
Net Power Capacity From Coal	MW	255	595
Net Power Capacity From Biomass	MW	45	105
Annual Electricity From Coal	GWh/yr	2,136	4,984
Annual Electricity From Biomass	GWh/yr	377	880
Coal Consumption (wet)	tonnes/yr	802,220	1,871,847
Biomass Consumption (Dry)	tonnes/yr	218,347	509,476
Annual Heat Input From Coal	TJ/yr	23,638	55,155
Annual Heat Input From Biomass	TJ/yr	4,172	9,735
Total Annual Heat Input	TJ/yr	27,810	64,890

Table 5.10: Biomass Cofiring Capital and Operating Costs, Excluding Feed

	1996$	2001$			
	300MW	300 MW	300 MW	700 MW	
Capital Cost (basis: Biomass Pow er Capacity)	$/kW	$/kW	$	$/kW	$
M&S Index	1039	1092			
Biomass Handling System Equipment					
Conveyor	10.3	10.8	487,097	8.4	881,452
Separation Equipment, Conveyor	2.8	2.9	132,415	2.3	239,618
Hogging Tow er and Equipment	17.0	17.9	803,946	13.9	1,454,824
Pneumatic Conveying System (Vacuum)	3.6	3.8	170,247	2.9	308,080
Wood Silo w ith Live Bottom	4.4	4.6	208,080	3.6	376,543
Collecting Conveyers	5.3	5.6	250,642	4.3	453,563
Rotary Airlock Feeders	0.5	0.5	23,645	0.4	42,789
Pneumatic Conveying System (Pressure)	13.6	14.3	643,157	11.1	1,163,859
Controls	8.4	8.8	397,244	6.8	718,854
Total Equipment	65.9	69.3	3,116,472	53.7	5,639,583
Installation	40.9	43.0	1,934,199	33.3	3,500,136
Total Biomass Handling	106.8	112.2	5,050,671	87.0	9,139,718
Civil Structure Work	29.4	30.9	1,390,353	24.0	2,515,990
Modifications at Burners	2.4	2.5	113,498	2.0	205,387
Electrical	13.1	13.8	619,511	10.7	1,121,070
Subtotal (A)	151.7	159.4	7,174,033	123.6	12,982,165
Contingency @ 15%, 0.15*A	22.8	23.9	1,076,105	18.5	1,947,325
Total Direct Costs (B)	174.5	183.3	8,250,138	142.2	14,929,490
Engineering @ 10%, 0.1*B	17.4	18.3	825,014	14.2	1,492,949
Total Capital Requirements (TCR)	191.9	201.7	9,075,152	156.4	16,422,439
Incremental Operating and Maintenance Costs					
Variable Costs					
Consumables	0.00163	0.0016	614,510	0.0016	1,433,857
Coal Savings		(0.0051)	(1,916,829)	(0.0051)	(4,472,601)
Fixed Costs					
Labor	0.0006	0.00063		0.00063	
Maintenance	0.0005	0.00053		0.00053	
Total	0.0011	0.00116	435,812	0.00116	1,016,895

Table 5.11: Cofiring Cogeneration Gross Steam Turbine Efficiencies, 100 MW_{eq} Plant

	Temp		Pressure		Flow		Quality	Electricity	Steam	HP Steam	Reheat	Efficiency		
	(°C)	(°F)	(MPa)	(psia)	(kg/s)	(lb/hr)	(%)	(MW_e)	(MW_t)	(MW_t)	(MW_t)	Electric (%)	Thermal (%)	Steam (%)
Condensing Turbine														
Inlet	538	1,000	16.50	2,048	95.51	758,046	100	106	0	303	39	36.8	0	36.8
Extraction[a]														
Outlet	54	130	0.0152	2.2	95.51	758,046	87							
Backpressure Turbine														
Inlet	538	1,000	16.50	2,048	95.51	758,046	100	63	260	303	39	17.5	76.6	94.1
Extraction[a]														
Outlet	212	414	1.140	165	95.51	758,046	100							
Extraction Turbine														
Inlet	538	1,000	5.860	865	95.51	758,046	100	94.5	130	303	39	26.1	38.3	64.4
Extraction[a]	212	414	1.140	165	47.76	379,023	100							
Outlet	54	130	0.0152	2.2	47.76	379,023	87							

[a] Doesn't include Process Extraction

Case Studies

A technoeconomic comparison has been made the direct combustion, gasification, and cofiring systems. A listing of cases, along with technical performance is given in Table 5.12. Plants are defined in terms of electricity-only base cases. For example, the 25 MW CHP case has a feed rate equal to the feed rate for a 25 MW_e electricity-only plant. The actual electric capacity for the 25 MW CHP case is 19.8 MW_e, and the plant also produces 107,000 lb/hr of 150 lb steam. On an energy basis, the H/P ratio is 1.44; and the overall HHV efficiency is 62%.

Table 5.12: Biomass Plant Technical Performance

Case	Efficiency %	Feed Rate MBtu/hr (TPH*)	Electricity MW	150 lb Steam 1000 lb/hr	H/P
25 MW Electric - Direct Comb	30	284 (16.73)	25.0	0	--
25 MW CHP - Direct Comb	62	284 (16.73)	19.8	107	1.44
25 MW Steam	75	284 (16.73)	-2.5	214	--
50 MW Electric - Direct Comb	30	569 (33.45)	50.0	0	--
50 MW CHP - Direct Comb	62	569 (33.45)	41.5	214	1.44
50 MW Steam	75	569 (33.45)	-5.0	429	--
75 MW Electric - Direct Comb	30	853 (50.18)	75.0	0	--
75 MW CHP - Direct Comb	62	853 (50.18)	62.2	321	1.44
75 MW Steam	75	853 (50.18)	-7.5	643	--
100 MW Electric - Direct Comb	30	1,137 (66.90)	100.0	0	--
100 MW CHP - Direct Comb	61	1,137 (66.90)	83.0	428	1.44
100 MW Steam	75	1,137 (66.90)	-10.0	857	--
75 MW Gasification-Electric	36	711 (41.80)	75.0	0	--
75 MW Gasification - CHP	82	711 (41.80)	59.3	324	1.60
150 MW Gasification - Electric	36	711 (41.80)	150.0	0	--
150 MW Gasification - CHP	82	1,422 (83.60)	118.6	648	1.60
45 MW Cofiring CHP (15%)	60	518 (30.46)	41.0	170	1.21
45 MW Cofiring Steam	66	518 (30.46)	-2.7	341	--
105 MW Cofiring CHP (15%)	60	1,208 (71.08)	95.7	397	1.21
105 MW Cofiring Steam	66	1,208 (71.08)	-6.30	796	--

* Dry tons @ 17 MBtu/ton

For each of the cases a discounted cash flow (DCF) analysis was performed, using the economic parameters presented in Table 5.1. Since CHP operations have two products, electricity and steam, a protocol for prorating values was needed. One way to do this would be to assign market value to one product and determine the required cost of the second. However, this can unduly penalize or benefit the product being calculated if the required cost differs significantly from market value. Therefore, this method was not used. A second method is to estimate the present market value of the two products, and use the ratio to determine required costs of both. An estimate of relative market values was made using EIA[2] cost of manufacturing data from 1998. Based on survey data, the EIA presented purchased electricity

[2]Energy Information Agency, Department of Energy, EIA Manufacturing Consumption of Energy 1998, Table N8.3, 1998.

and steam data for the United States and census region by manufacturing sector. A complete set of cost data for chemical industry sub-sectors in given in Appendix 2. For this study, the average values for the chemical sector were used. Figure 5-7 shows census regions. Figures 5-8 and 5-9 show purchased electricity and steam costs, respectively, updated to 2001 dollars using the GDP deflator. The United States average value of industrial electricity was $0.038/ kWh and industrial steam was $3.20/1,000 lb. In practice, as can be seen in the figures, the actual ratio will be site specific. When converted to a consistent set of units, the ratio of heat to power value ($H/$P) was 0.287. The matrix of cases analyzed was relatively large. For example, for direct combustion there were four plant sizes and five feed cost levels. A graphical presentation of the results using the 0.287 ratio with electricity in ¢/kWh and steam in $/1,000 lb was confusing simply because of the number of lines on each graph. In discounted cash flow analysis, if the capital and operating costs are fixed, and the discount rate is held constant, all feasible solutions give identical cash flows, e.g. identical incomes. Therefore, the absolute ratio of electricity and steam costs does not significantly impact the analysis. A $H/$P value was determined, 0.341, which would simplify graphical presentation of results, and would still be realistic; this value was used for the case studies.

Figure 5.7: United States Census Regions

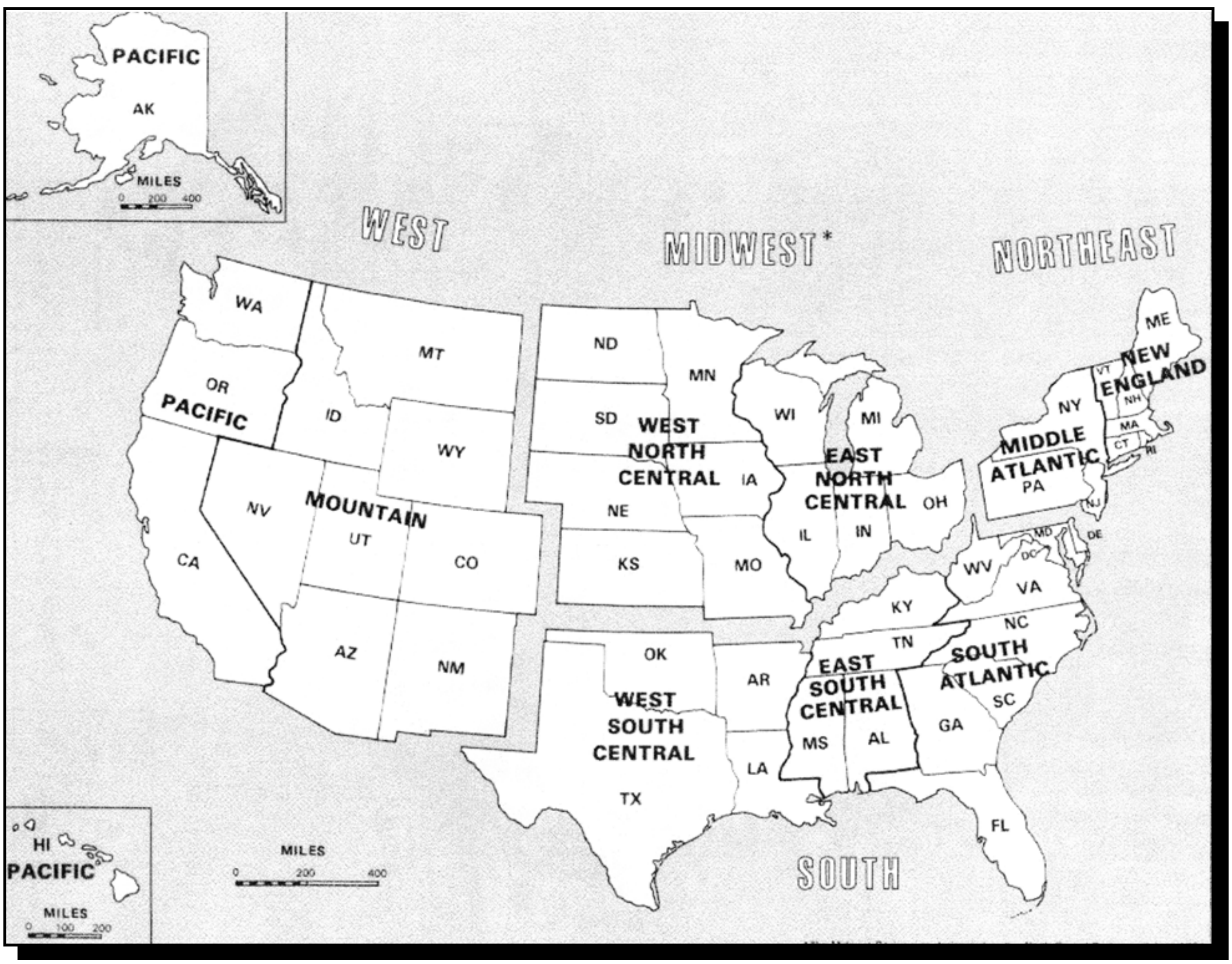
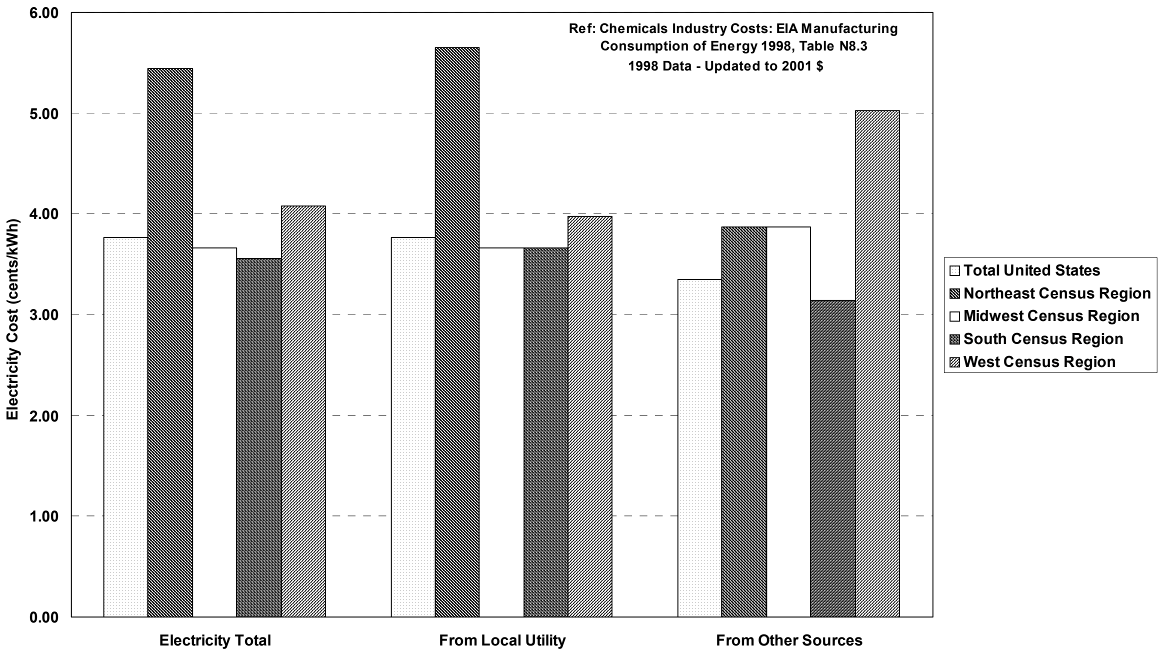


Figure 5.8: Chemical Industry Cost of Purchased Electricity



The DCF analysis was performed as a net present value (NPV) equals zero calculation in which the internal rate of return was set at the assumed discount rate, and the cost of products varied until NPV equaled zero. An example set of input values is given in Table 5.13, and the corresponding cash flow result shown in Table 5.14.

Feed Cost

In Figure 5.10, the effect of feed cost on required electricity and steam costs is shown for all systems The negative feed cost represents residue material generated in a chemical manufacturing or other industrial facility that is presently disposed of at some net cost, and where the negative cost represents a savings in disposal cost that can be represented by a negative transfer price. The 0 - 1 $/ton values represent residue materials presently used (see Figure 3.1 in Chapter 3), and the higher values represent marginal costs for larger supply levels. Typically, dedicated feeds will only be available at > $3/MBtu. The results show that all combustion CHP cases give required product costs greater than existing industrial market prices. The gasification plants show a comparable trend, but with required product costs 2 to 3¢/kWh ($2-3/1,000 lb steam) lower than the direct combustion cases. Gasification CHP using technology presently available, i.e., 1st generation commercial systems, may be competitive with existing sources of industrial electricity and steam if a manufacturing facility has an internal source of waste available. For higher cost residues or dedicated crops, incentives or more advanced technologies, i.e., nth plant technology with higher efficiency, will be required to reduce product costs to a competitive level. Cofiring represents fuel substitution for existing coal feed. The coal savings offsets the required capital investment and the incremental cost of cofiring reflects the cost of biomass feed.

Plant Size

Figure 5.11 shows the effect of plant size on required product cost for the base feed cost of $2/MBtu. Capital and operating costs were scaled using a 0.7 scaling factor. The rationale for the scaling factor was discussed earlier in the direct-fired biomass section. Since only two plant sizes were calculated for gasification and cofiring, the shape of the curve is not apparent, but would follow the same trend if more sizes were estimated. The cost of electricity (steam) for direct combustion varied from 10.6 ¢/kWh ($/1,000 lb steam) at 25 MW_{eq} to 8.4 ¢/kWh ($/1,000 lb steam) at 100 MW_{eq}. Gasification production costs were 6.7¢/kWh ($/1,000 lb steam) at 75 MW_{eq} and 6.1 ¢/kWh ($/1,000 lb steam) at 150 MW. For cofiring at 15% the incremental costs were 2.2 ¢/kWh ($/1000 lb steam) for 45 MW_{eq} biomass and 2.1 ¢/kWh ($/1,000 lb steam) for 105 MW_{eq} biomass.

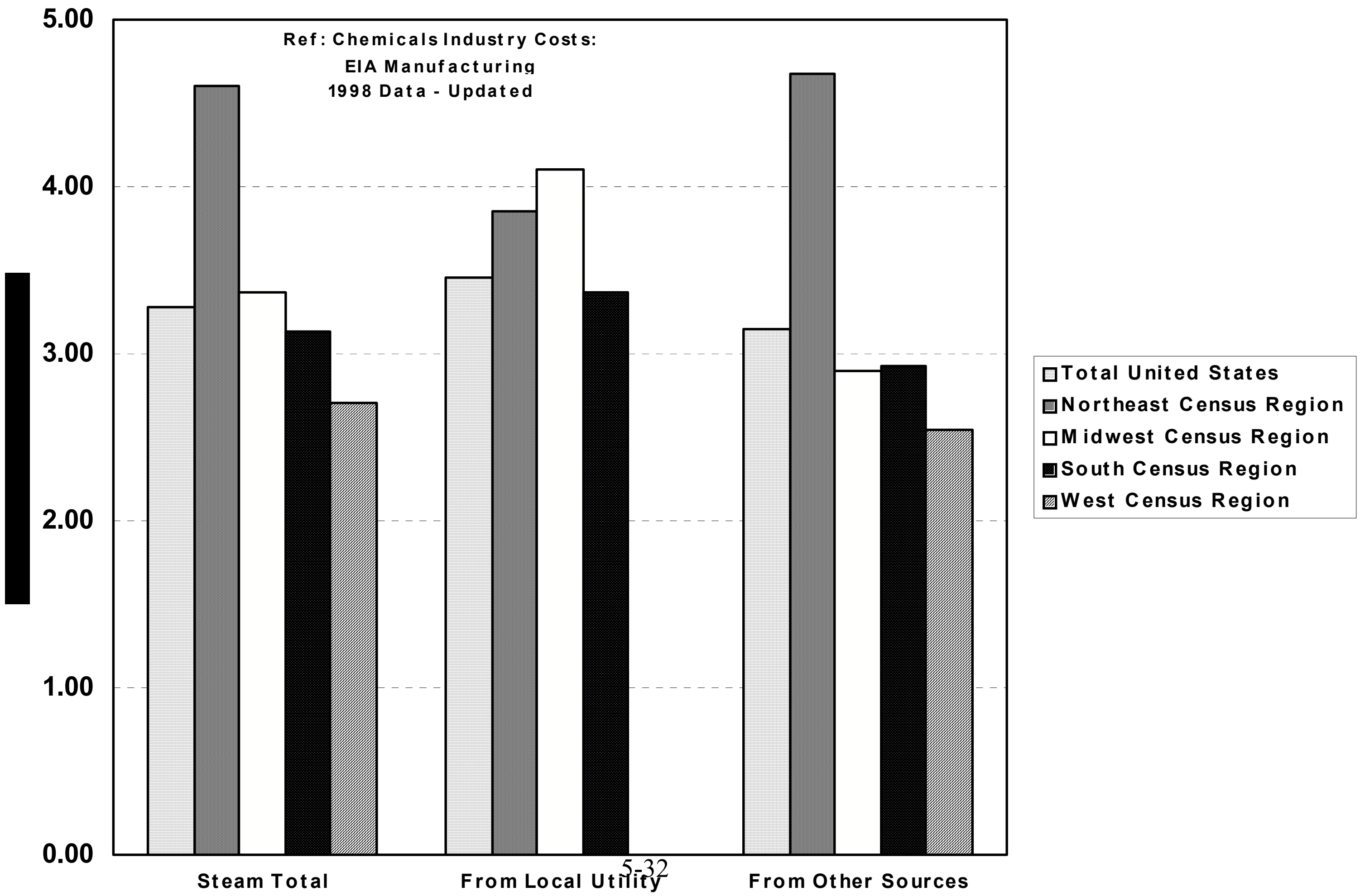
Capital Cost

The sensitivity of cost of production on capital cost is shown in Figure 5.12. Because of the low capital investment required for 15% cofiring CHP the sensitivity to a ± 25% variation in capital cost is small, e.g., 2.1 ± 0.08 ¢/kWh ($/1,000 lb steam) at 105 MW_{eq} biomass. For gasification CHP, the sensitivity at 150 MW_{eq} is 6.10 ± 0.73 ¢/kWh, and for direct combustion at 100 MW_{eq} is 8.44 ± 0.97 ¢/kWh for the same ± 25% variation.

Discount Rate

The base case study uses a 20 percent discount rate, but the allowable discount rate is dependent on the

Figure 5.9: Chemical Industry Cost of Purchased Steam



individual organization performing the evaluation. NREL typically uses 15% for analyses, and EPRI has used 10.8% for utility cash flow comparisons (EPRI TAG, 1993). Therefore, a set of sensitivity cases was performed to look at the sensitivity to discount rate over the range 10%-25%. The results are given in Figure 5.13. Over the range of 10% to 25 %, the 100 MW_{eq} direct combustion CHP system cost varies from 6.4 ¢/kWh ($/1,000 lb steam) to 9.6 ¢/kWh, the 150 MW_{eq} gasification CHP cost from 4.5¢/kWh to 7.0¢/kWh, and the 105 MW_{eq} cofiring CHP cost from 1.9¢/kWh to 2.2¢/kWh.

Debt Sensitivity

Each organization has its own debt-equity protocol for cash flow estimation. A set of sensitivity cases was performed to estimate the required cost sensitivity to level of debt and the results are shown in Figure 5.14. For the 100 MW_{eq} direct combustion system, the required CHP costs are 10.5¢/kWh ($/1,000 lb steam) and 7.5¢ at 100% equity and 50%equity-50% debt, respectively. For the 150 MW_{eq} gasification system the respective CHP costs at 100% and 50% equity are 7.6¢/kWh and 5.4¢/kWh, and for cofiring CHP at 105 MW_{eq} are 2.2 and 2.0 ¢/kWh.

Carbon Allowances

There is a significant amount of ongoing discussion about the use of carbon taxes, carbon sequestration credits, or carbon emission penalties as a way to reduce greenhouse gas emissions. An estimate of the impact of carbon allowances on biomass CHP was estimated over the range of 0-100 $/ton carbon emissions avoided. U.S. average carbon emissions for electricity production from coal-fired utility stations were used to estimate carbon allowances (EIA AEO 2002). The estimated value is 2.75×10^{-4} metric tons carbon/kWh. The carbon allowance was credited only against electricity production and was considered a before-tax income stream. The results are given in Figure 5.15. The cost of electricity (steam) for the 100 MW_{eq} direct combustion CHP case at 0, 25, 50, and 100 $/metric ton carbon avoided are 8.44, 8.01, 7.58, and 6.71 ¢/kWh ($/1,000 lb steam), respectively. For the 150 MW_{eq} gasification CHP case, the respective costs are 6.10, 5.65, 5.21, and 4.32 ¢/kWh; and for 15% cofiring CHP at 105 MW_{eq} biomass the costs are 2.06, 1.58, 1.09, and 0.12 ¢/kWh.

Tax Credits

Various proposals are before Congress to modify and expand the definition of the IRS Section 49 "closed loop" biomass tax credit to include residues and cofiring. Therefore, estimates of the impact of such tax credits were made. The estimates were made using two assumptions–a project basis and a corporate basis. Using the project basis assumption, only project generated taxable income is used. In this case, the capital equipment depreciation in the early years of the project greatly limits taxable income, and the impact of a tax credit is small. For the corporate basis cases, the assumption is made that the corporation has other taxable income that the tax credit can be applied against so that all potential tax credit can be used. Other assumptions are that the tax credit is available for 100% of the net plant production, i.e, that the net production of electricity is sold, and that the tax credit applies for 10 years of plant operation.

Figure 5.16 shows the impact of a electricity production tax credit on the required cost of production of electricity for direct combustion CHP and for gasification CHP. For the 100 MW_{eq} direct combustion system, the impact on electricity (steam) cost of production with a 1¢/kWh production credit is minus 0.36 ¢/kWh ($/1,000 lb steam) for a project basis and minus 0.77 ¢/kWh for a corporate basis; with a 2 ¢/kWh production credit the respective values are minus 0.57¢/kWh and minus 1.51 ¢/kWh. For the 150 MW_{eq} gasification CHP system with a 1¢/kWh production credit, the cost of production is lowered by 0.42 and 0.81 ¢/kWh for the project and corporate cases, respectively; with a 2¢/kWh credit, the respective cost of production reduction is 0.50 and 1.57 ¢/kWh.

Comparable estimates can be made for the cofiring CHP cases, but the analysis is somewhat more complicated. Figure 5.17 shows the impact of a production credit on 15% cofiring CHP incremental costs. For the project cases, the decrease in cost of production for the 45 and 105 MW_{eq} plants reaches a maximum of about 0.07 ¢/kWh at a tax credit of about 0.5¢ /kWh. For the corporate analysis, the NPV calculation does not give meaningful results above a tax credit level of 0.5¢/kWh. At this level, the reduction is about 0.40 ¢/kWh for both plants sizes. Above this level a NPV calculation can be made but to satisfy the 20% return constraint, a solution is obtained that gives negative cash flows in the years after expiration of the tax credit. This indicates that the 10-year production tax credit has a large impact on potential project rate of return. A return on investment (ROI) estimate was made to show the impact of the production tax credit, assuming a fixed cost of production. For this example, the incremental cost of production of electricity (steam) was set at 2.0 ¢/kWh ($2/1,000 lb steam), and the 10-year production tax credit varied from 0 to 1 ¢/kWh. For the 45 MW_{eq} plant, the ROI varies from 13.6%, to 31.6%, to 47.8% at a 0, 0.5, and 1 ¢/kWh tax credit, respectively. The comparable ROIs for the 105 MW_{eq} case are 17.0, 38.6, and 57.5%.

Steam Only

For the direct combustion and cofiring systems based on the Rankine cycle, cases were analyzed to see if steam-only production was more economic. For the direct combustion system, the steam turbine was removed from the capital cost estimate, and for both systems the steam exiting the boiler was used as product. The results are shown in Figure 5.19. Production of steam results in an increase in the cost of production in comparison to CHP. At 50 MW_{eq} the cost of steam is 11.90 $/1,000lb, an increase of 2.53 $/1,000 lb over the comparable CHP case. At 100 MW_{eq}, the steam cost is 10.81 $/1000lb, an increase of 2.37 $/1,000 lb. For cofiring steam-only production at 45 MW_{eq} the incremental cost of steam is 4.16 $/1000 lb, versus 2.16$/1000 lb for CHP; at 105 MW_{eq}, the incremental cost of steam is 3.82 $/1,000 lb versus 2.06 $/1,000 lb for CHP.

Site and Incentives Impact

To see the potential impact of regional plant location, discount rate, and incentives, one set of gasification CHP cases was performed. The discount rate was assumed to be 5%, and both a carbon allowance (15$/ton carbon equivalent) and production tax incentive (1.5 ¢/kWh) were allowed. The tax incentive was taken on a corporate basis. The results were compared to U.S. and Northeast region costs of purchased electricity and steam, and are presented in Figure 5.20. For 75 MW_{eq} both the cost of electricity and steam are higher than the national average purchased prices, but are lower than purchase prices in the Northeast. At 150 MW_{eq} the results are comparable, but the cost of electricity is equal to the national average price.

Capital Requirements and Required Cash Flow

Although the three technologies–direct combustion, gasification, and cofiring–were evaluated at a constant discount rate to determine the required costs of electricity and steam, another important investment consideration is capital required and cumulative cash flow realized over the life of the project.
A comparison of costs and cash flow is given in Table 5.15. Cofiring has the smallest capital investment and lowest operating costs, and gives the smallest cash flow. Gasification at 150 MW has an intermediate investment requirement, and direct combustion has the highest requirements.

Recommendations for Further Work

The case studies give a good base comparison of the three technologies. Further analysis is needed to fully investigate CHP applications.

- An advanced gasification case with higher efficiency-to-electricity ratios should be analyzed to determine if costs can be reduced to purchase prices.

- Given the variation in regional prices of electricity and steam, a resource evaluation relative to chemical plant locations should be performed to determine if there are site specific cases where feasibility studies should be performed. The project should map currently available biomass feedstocks (e.g., industrial processing residues, urban wood residues, agricultural residues) against locations of industrial facilities capable of biorefinery operations. The mapping should allow for a preliminary identification and ranking of prospective opportunities based on feedstock characteristics (i.e., type and availability, processing requirements, delivered costs) and industrial facility characteristics (i.e., type, size, location). It should utilize a biomass feedstock database being developed by ORNL that provides quantity and delivered price information at the county level. Feedstock evaluations should include secondary residues generated by biorefineries. It should also utilize secondary data and information sources, such as EPA Sector Notebooks and U.S. Census Bureau County Business Patterns, where appropriate.

- For a few of the more promising identified facilities, we should apply the technical and economic findings of this report to determine the appropriate biomass CHP systems (e.g., gasification, co-firing) as well as syngas opportunities for chemicals, such as ethanol or mixed alcohols. This latter analysis should assess opportunities for specific facilities as well as develop a replicable methodology (including information requirements) for identifying biorefinery opportunities for industrial sites throughout the U.S.

Table 5.13: Cash Flow Analysis Input Data, 75MW$_{eq}$ Gasification CHP System

Construction Period			2	yrs					
Operating Life			30	yrs					
Electrlcity Prod	4.6750E+08	kWh/yr							
Electricity Rate	0.06711	$/kWh							
Capital Cost	115,991,000								
Depreciable Capita	110,001,000								
3 year									
5 year									
7 year	82,454,000								
10 year									
15 year									
20 year	27,547,000								
% capital year 1	75%								
Inflated Capital	115,991,000								
Inflated depreciabl	110,001,000								
Variable Cost (I = 0)									
year 1									
year 2									
year 3-n	3,294,000								
Fixed Cost	3,421,000								
Feed									
ton/yr	329,847								
$/ton (I=0)	34.00								
$/yr	11,214,798								
Working Cap	22,500,000								
Revenues									
Electricity	31,374,189								
Capacity Paymt									
Coproduct	17,112,492								
Total Revenue	48,486,681								
Yr 3 Prod, %	75.0%								
Inflation rate, %	0.00%								
Percent Debt	50.0%								
Cost of Debt	7.0%								
Term	20								
Inflated Debt	57,995,500								
Payment	5,474,365								
Depreciation Schedule-MACRS w half-yr convention			3-yr		5-yr		7-yr	10-yr	15-yr
Percentage			0%		0%		75%	0%	0%
Taxes									
Federal			35.0%						
State			5.0%						
State Wholesale Excise			0.0%			Coal Carbon			2.75E-04
Carbon Allowance			0.000	$/kWh		Carbon Allowance			0.000
Fed Tax Credit			0.000	$/kWh					
Corporate Basis (Yes=1, No = 0)			0						

Table 5.14: Cash Flow, 75 MW_{eq} Gasification CHP System

Year	Income I	Production Credit PC	Equity A	Debt	Debt Service B	Remaining Capital	Principal Paymt P	Working Capital C	Fixed Op cost D	Variable Op cost E	Pre-Dep Income F = I-A-B-C-D-E
1			43,496,625	43,496,625	3,044,764						(46,541,389)
2			14,498,875	14,498,875	4,059,685	57,995,500					(18,558,560)
3	36,365,011	0			4,059,685	56,580,820	1,414,680	22,500,000	3,421,000	10,881,599	(4,497,273)
4	48,486,681	0			3,960,657	55,067,113	1,513,708		3,421,000	14,508,798	26,596,226
5	48,486,681	0			3,854,698	53,447,446	1,619,667		3,421,000	14,508,798	26,702,185
6	48,486,681	0			3,741,321	51,714,402	1,733,044		3,421,000	14,508,798	26,815,562
7	48,486,681	0			3,620,008	49,860,045	1,854,357		3,421,000	14,508,798	26,936,875
8	48,486,681	0			3,490,203	47,875,883	1,984,162		3,421,000	14,508,798	27,066,680
9	48,486,681	0			3,351,312	45,752,830	2,123,053		3,421,000	14,508,798	27,205,571
10	48,486,681	0			3,202,698	43,481,163	2,271,667		3,421,000	14,508,798	27,354,185
11	48,486,681	0			3,043,681	41,050,480	2,430,683		3,421,000	14,508,798	27,513,202
12	48,486,681	0			2,873,534	38,449,648	2,600,831		3,421,000	14,508,798	27,683,350
13	48,486,681	0			2,691,475	35,666,759	2,782,890		3,421,000	14,508,798	27,865,408
14	48,486,681	0			2,496,673	32,689,067	2,977,692		3,421,000	14,508,798	28,060,210
15	48,486,681	0			2,288,235	29,502,937	3,186,130		3,421,000	14,508,798	28,268,649
16	48,486,681	0			2,065,206	26,093,778	3,409,159		3,421,000	14,508,798	28,491,678
17	48,486,681	0			1,826,564	22,445,977	3,647,800		3,421,000	14,508,798	28,730,319
18	48,486,681	0			1,571,218	18,542,830	3,903,147		3,421,000	14,508,798	28,985,665
19	48,486,681	0			1,297,998	14,366,464	4,176,367		3,421,000	14,508,798	29,258,885
20	48,486,681	0			1,005,652	9,897,751	4,468,712		3,421,000	14,508,798	29,551,231
21	48,486,681	0			692,843	5,116,229	4,781,522		3,421,000	14,508,798	29,864,041
22	48,486,681	0			358,136	0	5,116,229		3,421,000	14,508,798	30,198,747
23	48,486,681	0							3,421,000	14,508,798	30,556,883
24	48,486,681	0							3,421,000	14,508,798	30,556,883
25	48,486,681	0							3,421,000	14,508,798	30,556,883
26	48,486,681	0							3,421,000	14,508,798	30,556,883
27	48,486,681	0							3,421,000	14,508,798	30,556,883
28	48,486,681	0							3,421,000	14,508,798	30,556,883
29	48,486,681	0							3,421,000	14,508,798	30,556,883
30	48,486,681	0							3,421,000	14,508,798	30,556,883
31	48,486,681	0							3,421,000	14,508,798	30,556,883
32	48,486,681	0						(22,500,000)	3,421,000	14,508,798	53,056,883
33											
34											
35											
36											
37											

COE	6.71	cents/kWh
PROD CRI	0.00	cents/kWh
TAX CREI	0.00	cents/kWh

Project Internal Rate of Return

Rate of Return Estimate	
	30.00%
Internal Rate of Return	
	20.00%
Desired Return	20.00%
NPV	0

Table 5.14 (cont.)

Deprec G	Pre-tax Calculation Income F - G	Tax	Federal Tax Credit	After-Tax Income	Corporate Tax Credit	Cash Flow	Cum Cash Flow
	(46,541,389)	0		(46,541,389)		(46,541,389)	(46,541,389)
	(18,558,560)	0		(18,558,560)		(18,558,560)	(65,099,949)
0	(4,497,273)	0	0	(4,497,273)	0	(5,911,952)	(71,011,901)
26,596,226	0	0	0	0	0	25,082,518	(45,929,383)
24,661,584	2,040,602	816,241	0	1,224,361	0	24,266,278	(21,663,105)
12,000,083	14,815,479	5,926,192	0	8,889,288	0	19,156,327	(2,506,779)
8,936,902	17,999,973	7,199,989	0	10,799,984	0	17,882,529	15,375,751
8,810,756	18,255,924	7,302,370	0	10,953,555	0	17,780,149	33,155,899
8,709,640	18,495,932	7,398,373	0	11,097,559	0	17,684,146	50,840,045
4,923,124	22,431,061	8,972,425	0	13,458,637	0	16,110,094	66,950,139
1,229,147	26,284,055	10,513,622	0	15,770,433	0	14,568,896	81,519,035
1,228,872	26,454,478	10,581,791	0	15,872,687	0	14,500,727	96,019,762
1,229,147	26,636,261	10,654,504		15,981,756		14,428,014	110,447,776
1,228,872	26,831,339	10,732,535		16,098,803		14,349,983	124,797,759
1,229,147	27,039,501	10,815,801		16,223,701		14,266,718	139,064,477
1,228,872	27,262,806	10,905,122		16,357,684		14,177,396	153,241,873
1,229,147	27,501,172	11,000,469		16,500,703		14,082,050	167,323,923
1,228,872	27,756,793	11,102,717		16,654,076		13,979,801	181,303,724
1,229,147	28,029,738	11,211,895		16,817,843		13,870,623	195,174,347
1,228,872	28,322,359	11,328,944		16,993,416		13,753,575	208,927,922
1,229,147	28,634,894	11,453,957		17,180,936		13,628,561	222,556,483
1,228,872	28,969,876	11,587,950		17,381,925		13,494,568	236,051,051
614,574	29,942,310	11,976,924		17,965,386		18,579,959	254,631,010
	30,556,883	12,222,753		18,334,130		18,334,130	272,965,140
	30,556,883	12,222,753		18,334,130		18,334,130	291,299,270
	30,556,883	12,222,753		18,334,130		18,334,130	309,633,400
	30,556,883	12,222,753		18,334,130		18,334,130	327,967,530
	30,556,883	12,222,753		18,334,130		18,334,130	346,301,660
	30,556,883	12,222,753		18,334,130		18,334,130	364,635,790
	30,556,883	12,222,753		18,334,130		18,334,130	382,969,920
	30,556,883	12,222,753		18,334,130		18,334,130	401,304,050
	53,056,883	21,222,753		31,834,130		31,834,130	433,138,180

Figure 5.10: Biomass CHP - Sensitivity to Feed Cost

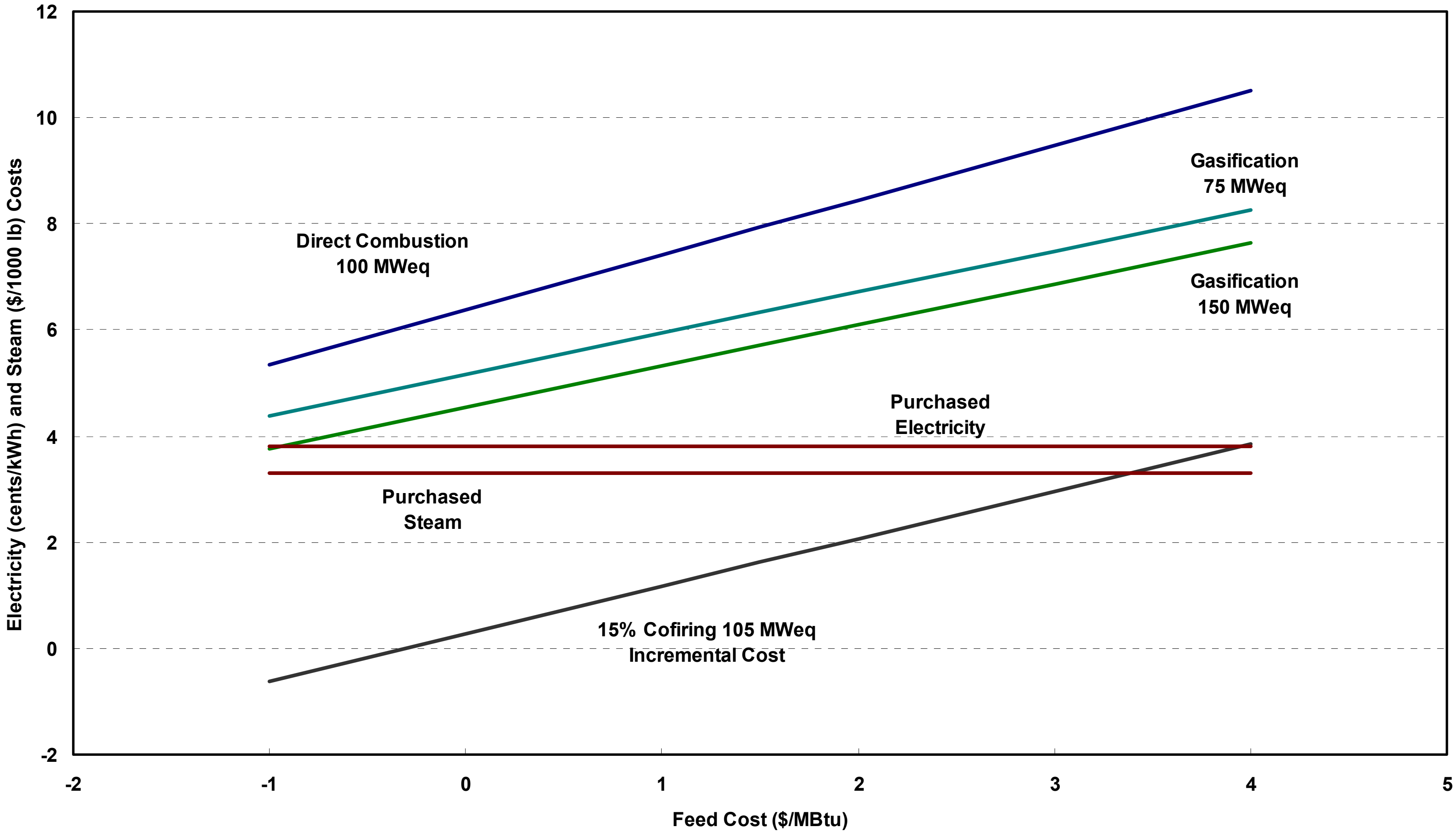
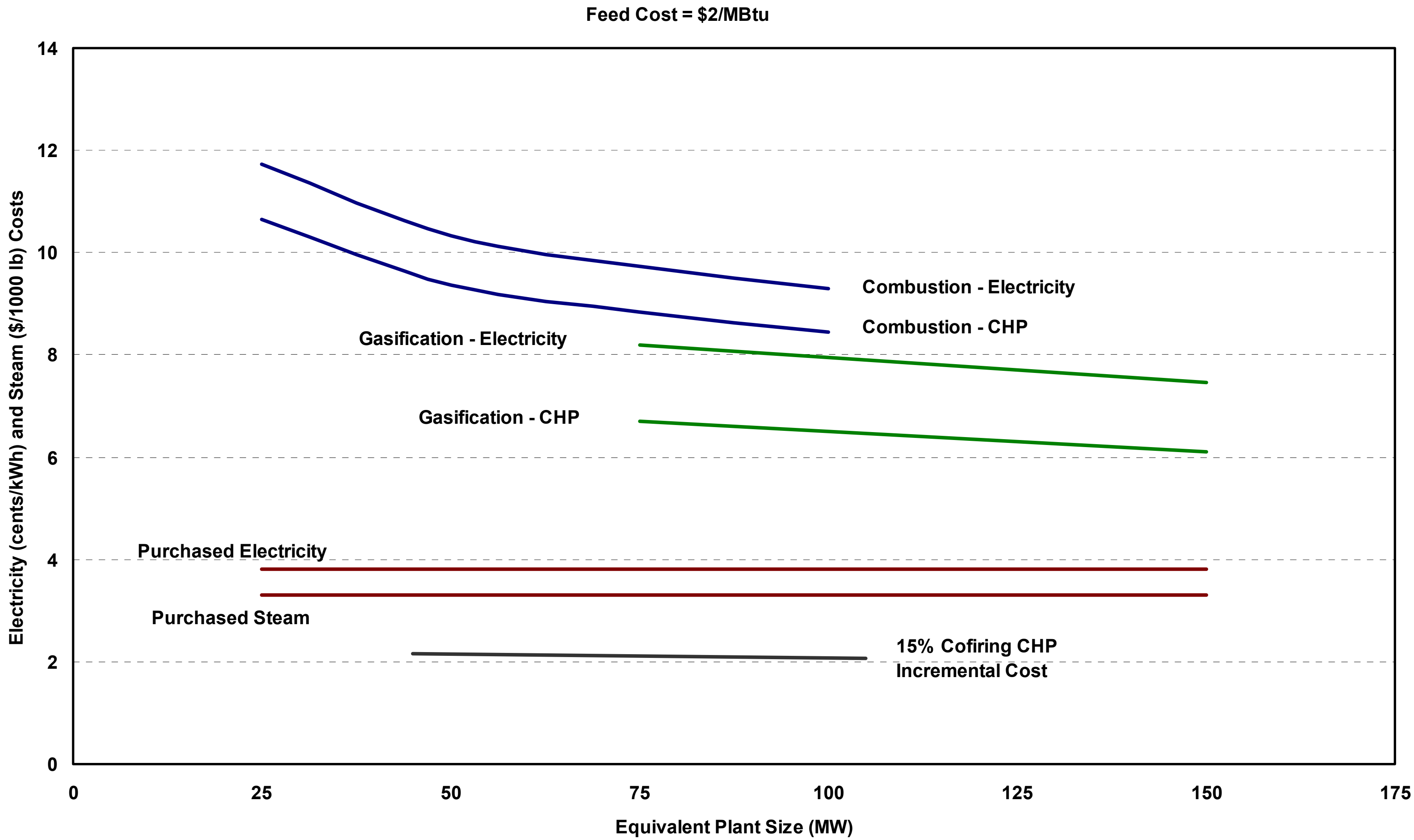


Figure 5.11: Biomass CHP - Effect of Plant Size on Cost of Electricity and Steam



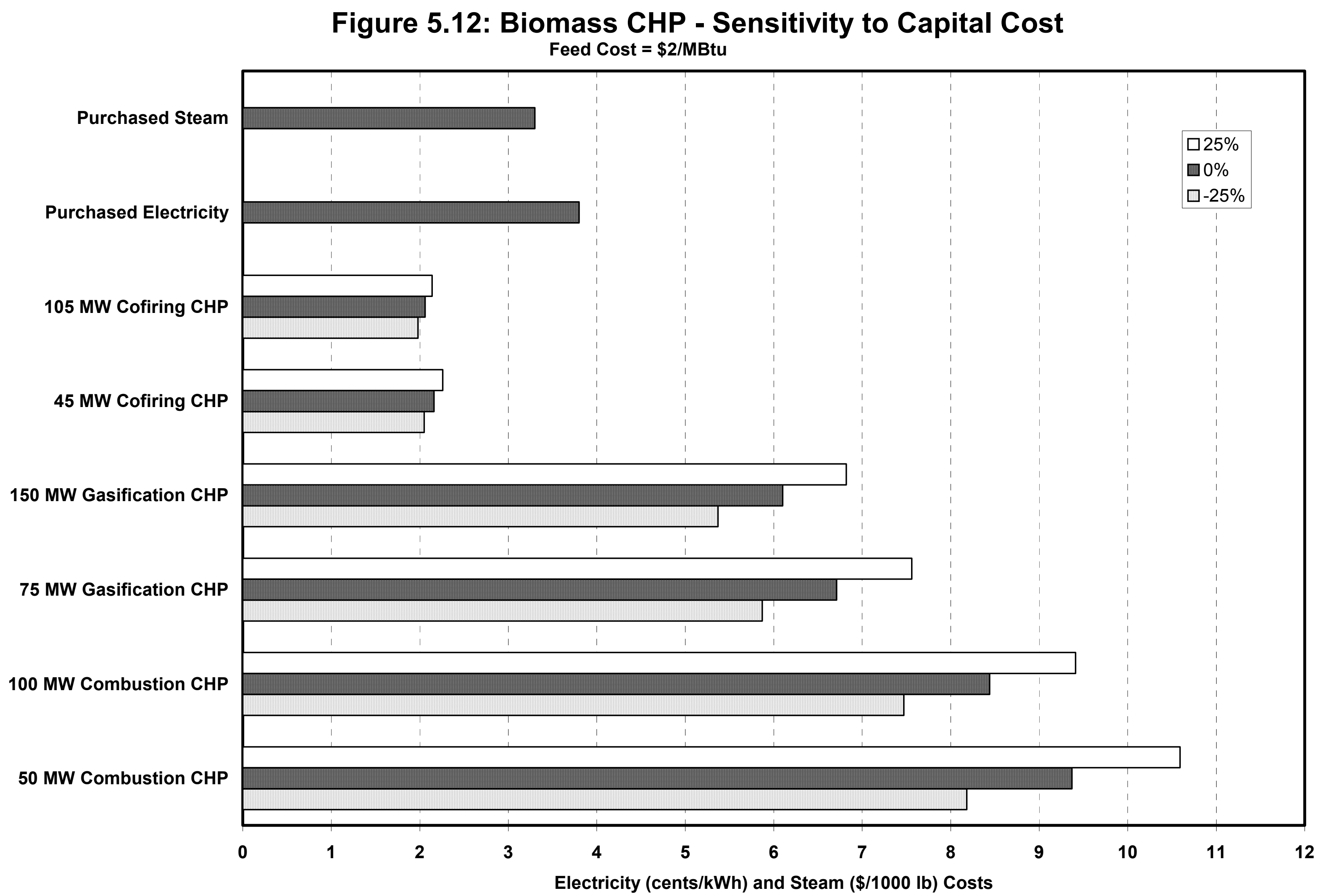
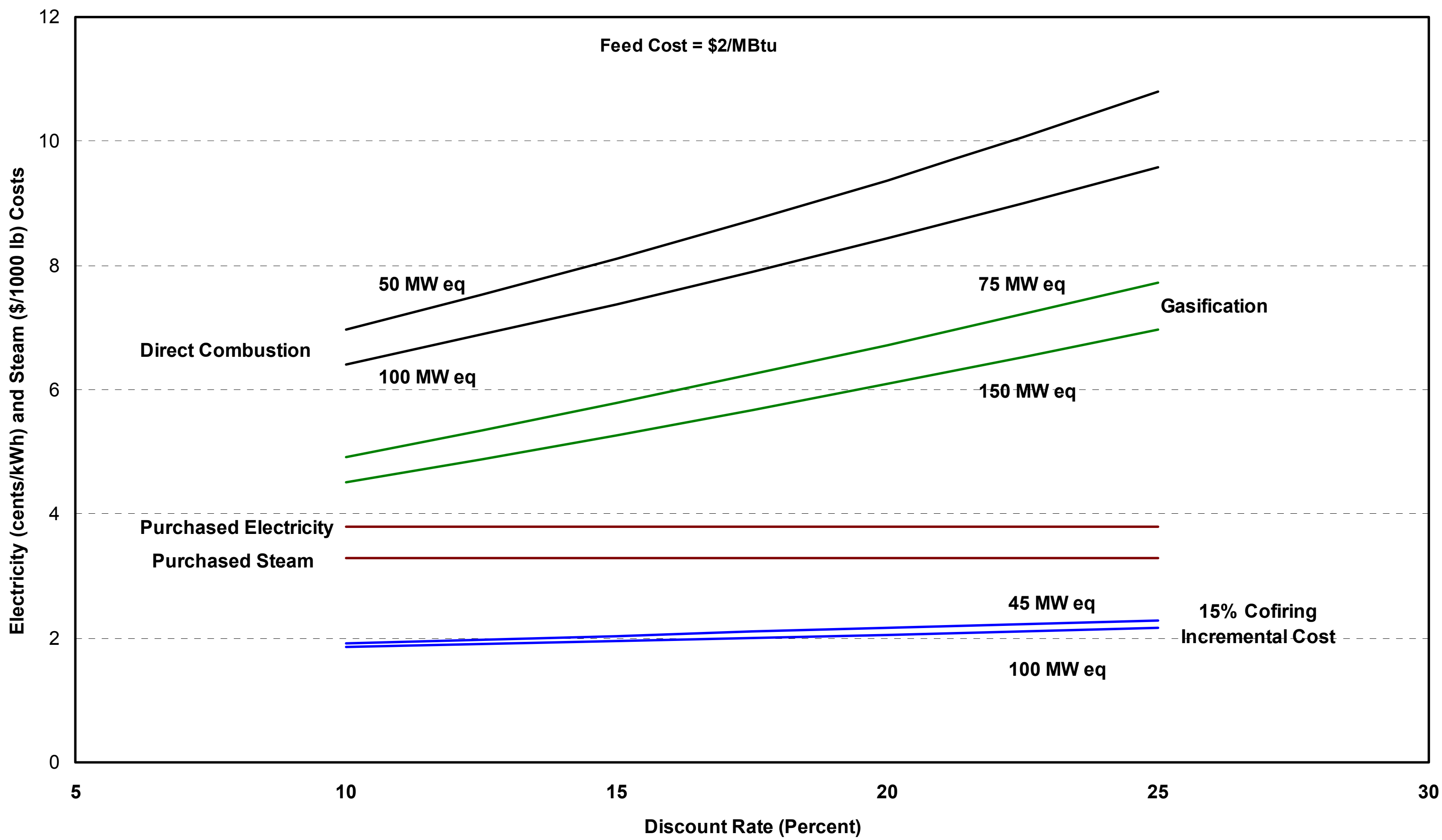


Figure 5.12: Biomass CHP - Sensitivity to Capital Cost
Feed Cost = $2/MBtu
Purchased Steam
Purchased Electricity
105 MW Cofiring CHP
45 MW Cofiring CHP
150 MW Gasification CHP
75 MW Gasification CHP
100 MW Combustion CHP
50 MW Combustion CHP
25%
0%
-25%
0
1
2
3
4
5
6
7
8
9
10
11
12
Electricity (cents/kWh) and Steam ($/1000 lb) Costs

Figure 5.13: Biomass CHP - Sensitivity to Discount Rate



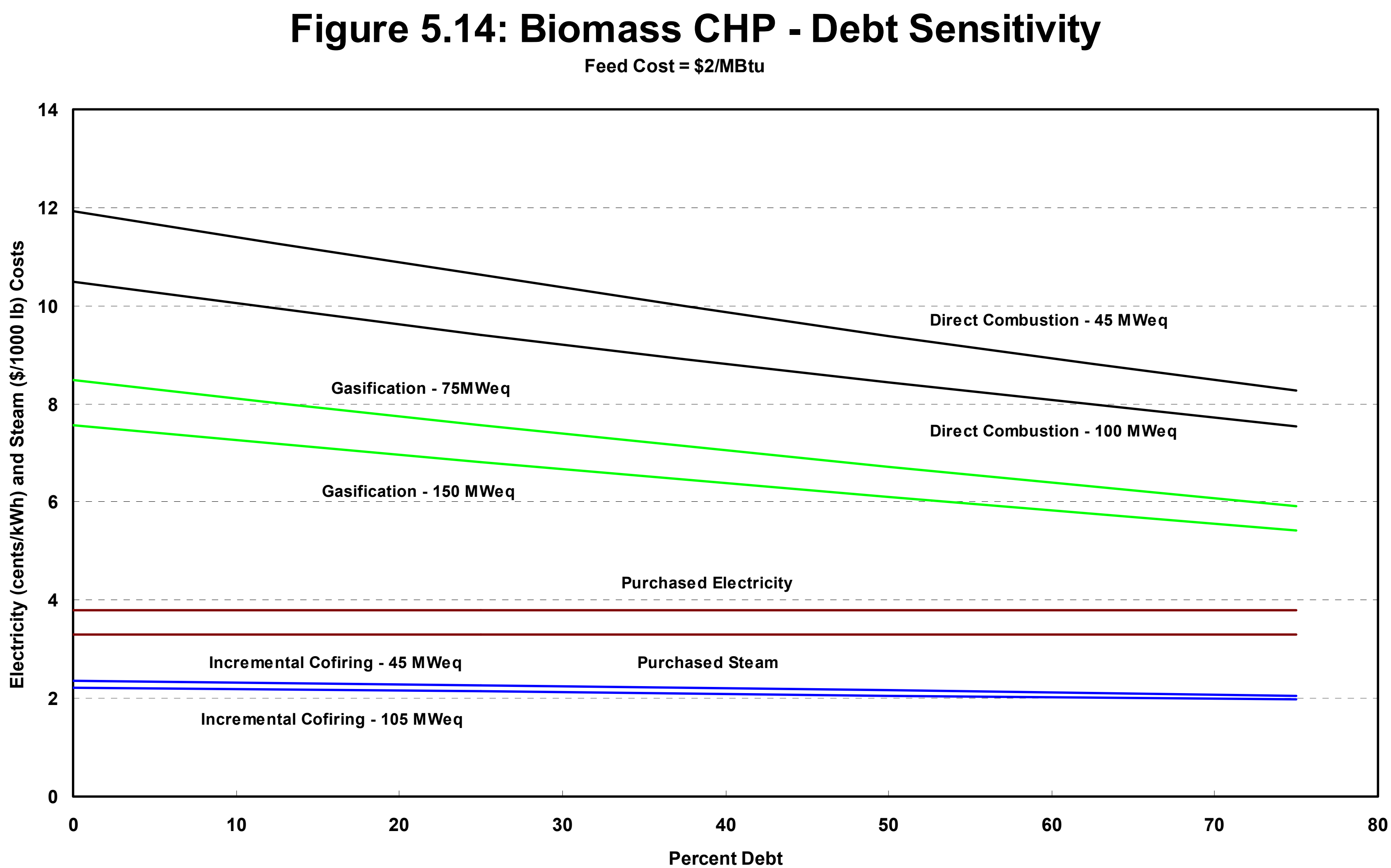


Figure 5.14: Biomass CHP - Debt Sensitivity
Feed Cost = $2/MBtu
Electricity (cents/kWh) and Steam ($/1000 lb) Costs
Percent Debt
Direct Combustion - 45 MWeq
Direct Combustion - 100 MWeq
Gasification - 75MWeq
Gasification - 150 MWeq
Purchased Electricity
Purchased Steam
Incremental Cofiring - 45 MWeq
Incremental Cofiring - 105 MWeq
0
2
4
6
8
10
12
14
0
10
20
30
40
50
60
70
80

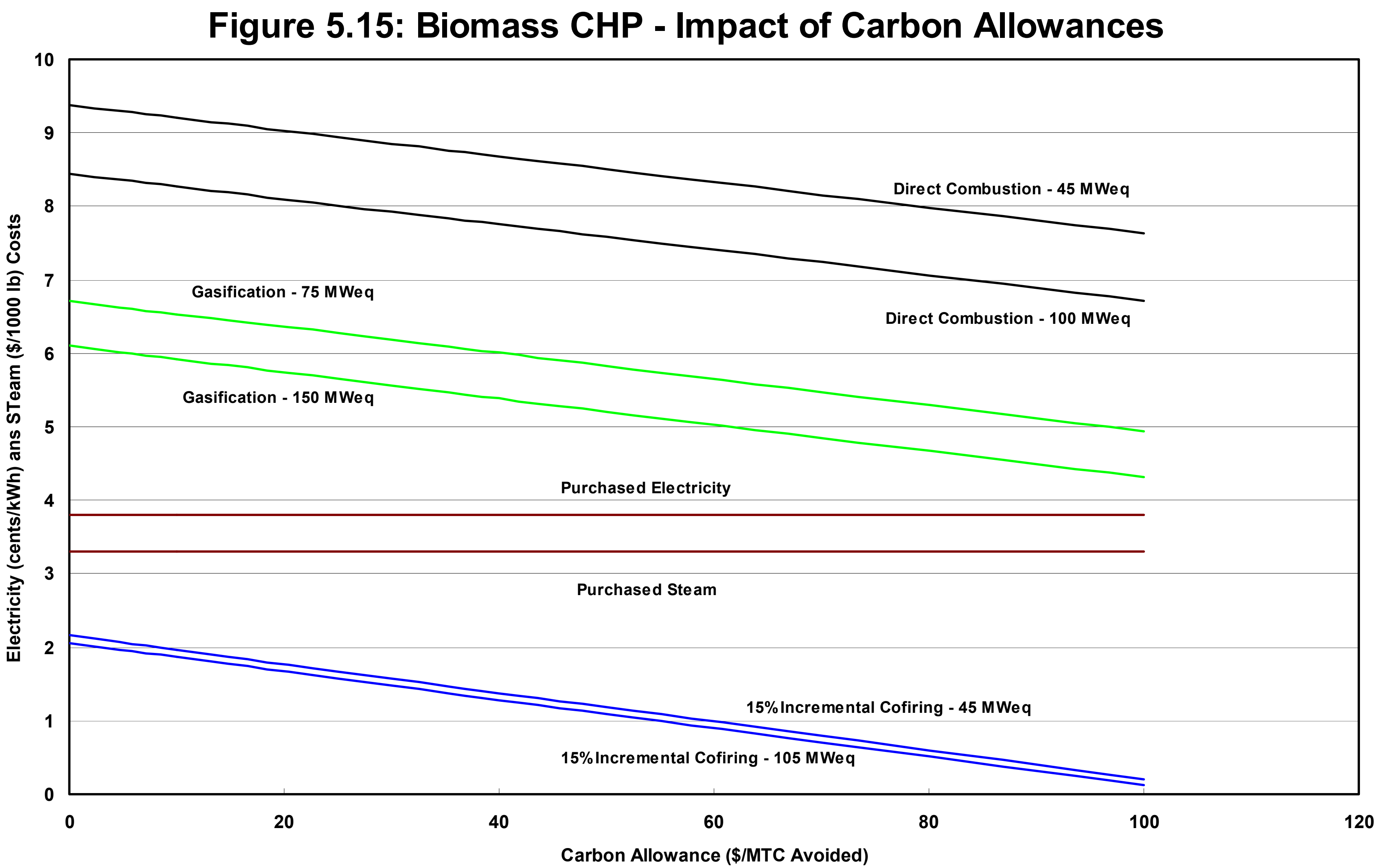


Figure 5.15: Biomass CHP - Impact of Carbon Allowances
Electricity (cents/kWh) ans STeam ($/1000 lb) Costs
Carbon Allowance ($/MTC Avoided)
Direct Combustion - 45 MWeq
Direct Combustion - 100 MWeq
Gasification - 75 MWeq
Gasification - 150 MWeq
Purchased Electricity
Purchased Steam
15% Incremental Cofiring - 45 MWeq
15% Incremental Cofiring - 105 MWeq
0
1
2
3
4
5
6
7
8
9
10
0
20
40
60
80
100
120

Figure 5.16: Biomass Combustion and Gasification CHP Impact of Tax Credit

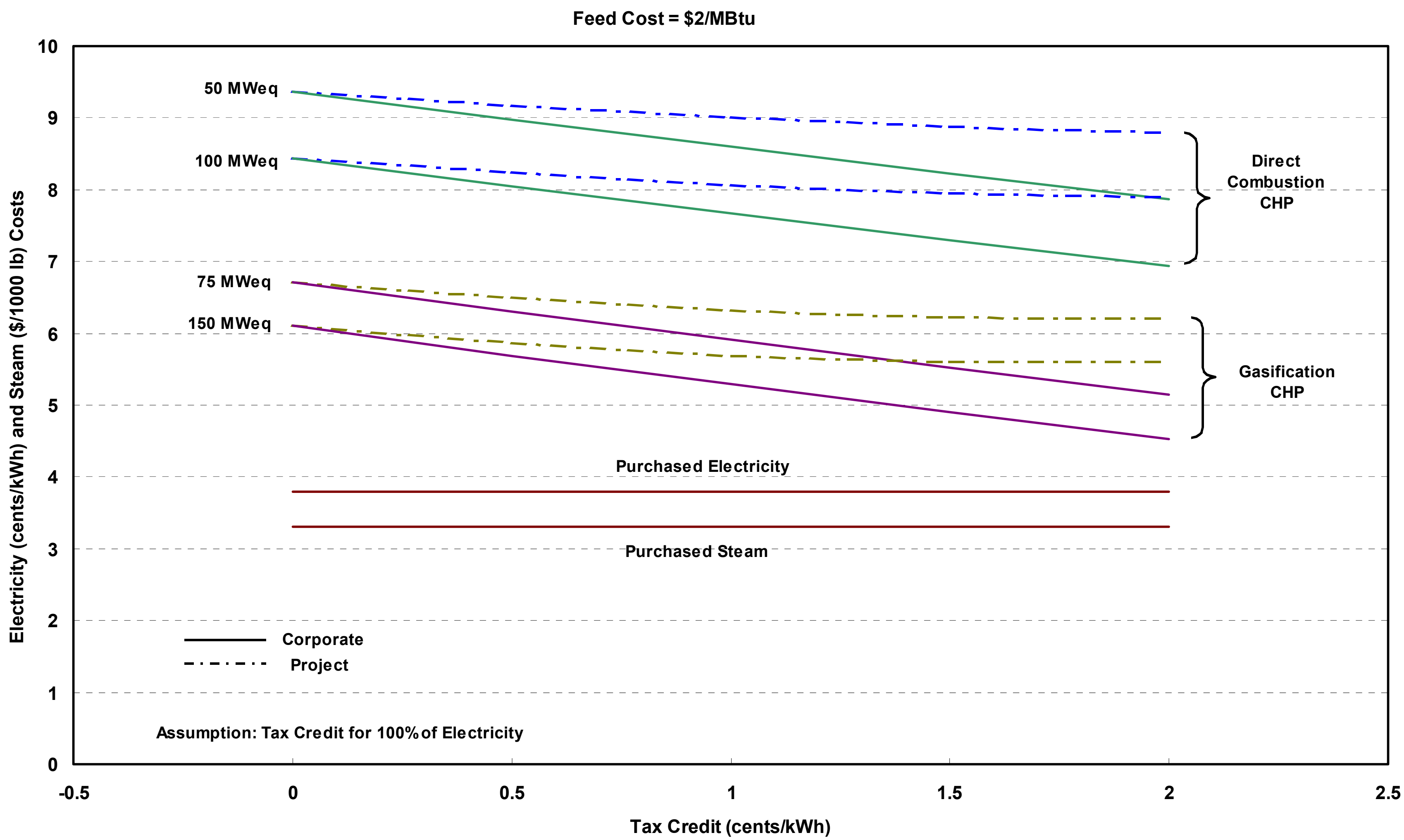


Figure 5.17: Biomass Cofiring CHP Incremental Costs, Impact of Tax Credit

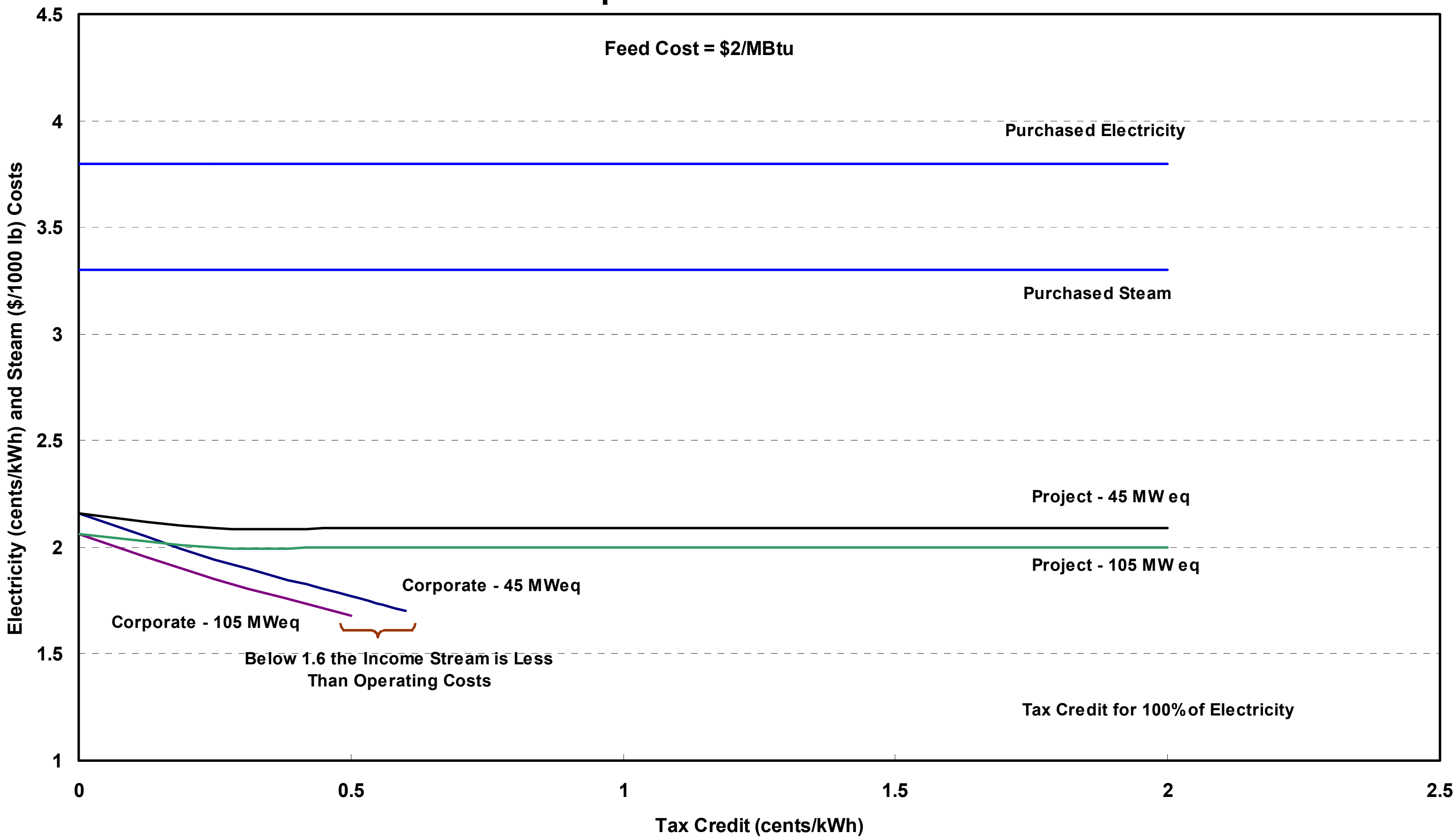


Figure 5.18: Biomass Cofiring CHP - Effect of Tax Credit on Return on Investment, Corporate Basis

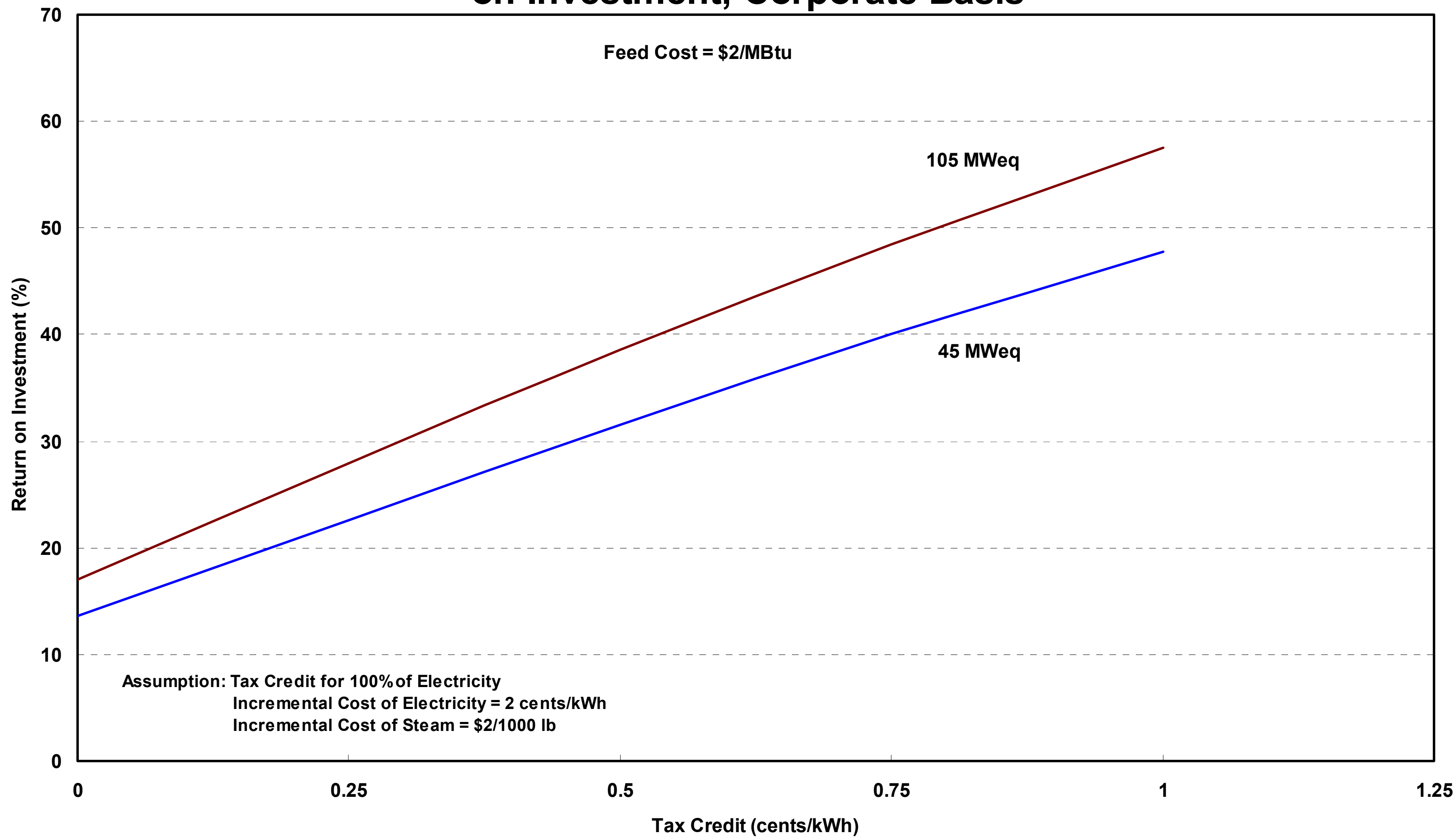


Figure 5.19: Comparison of CHP and Steam-Only Costs

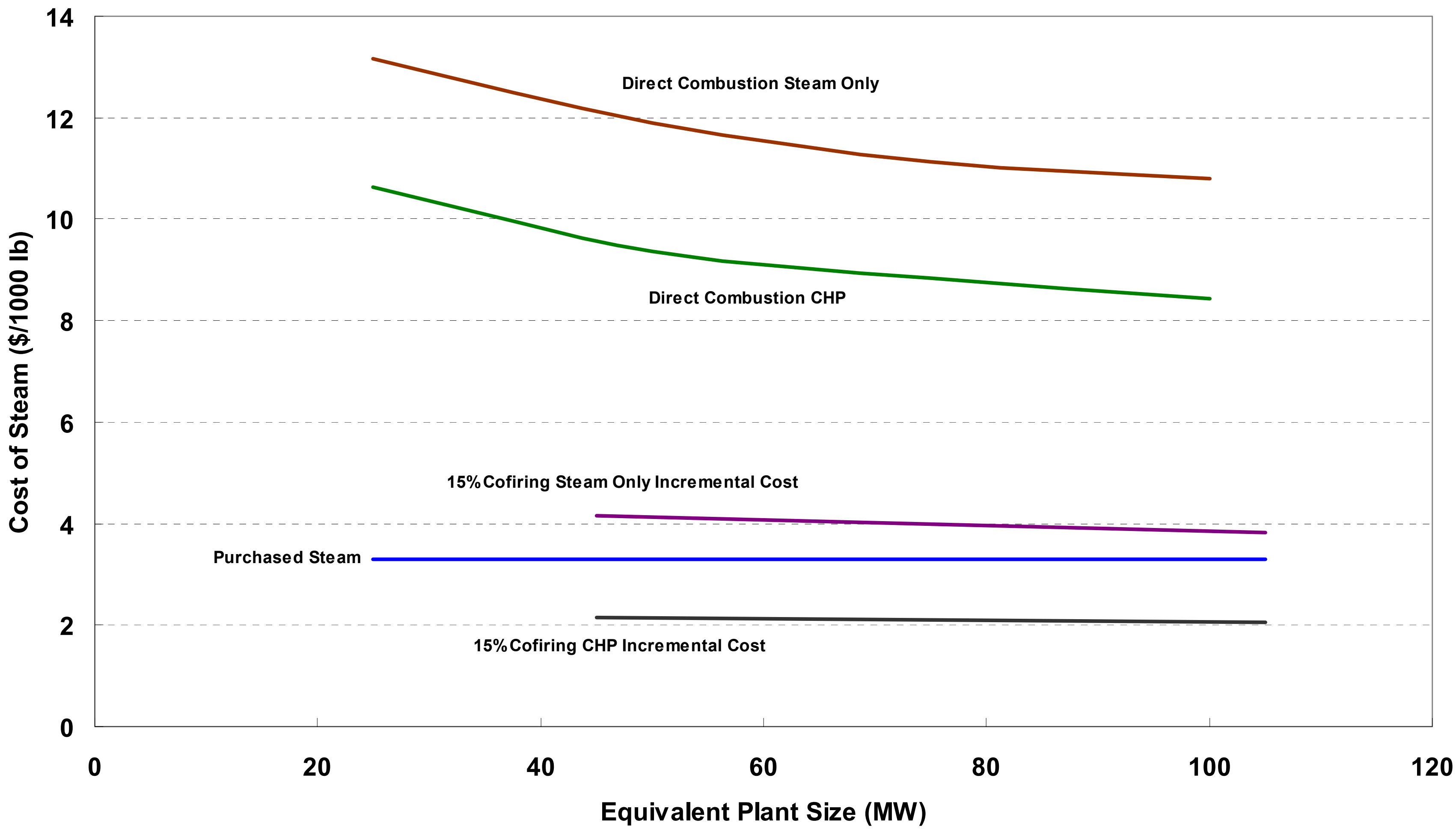


Figure 5.20: Gasification CHP - Site Impact with Incentives

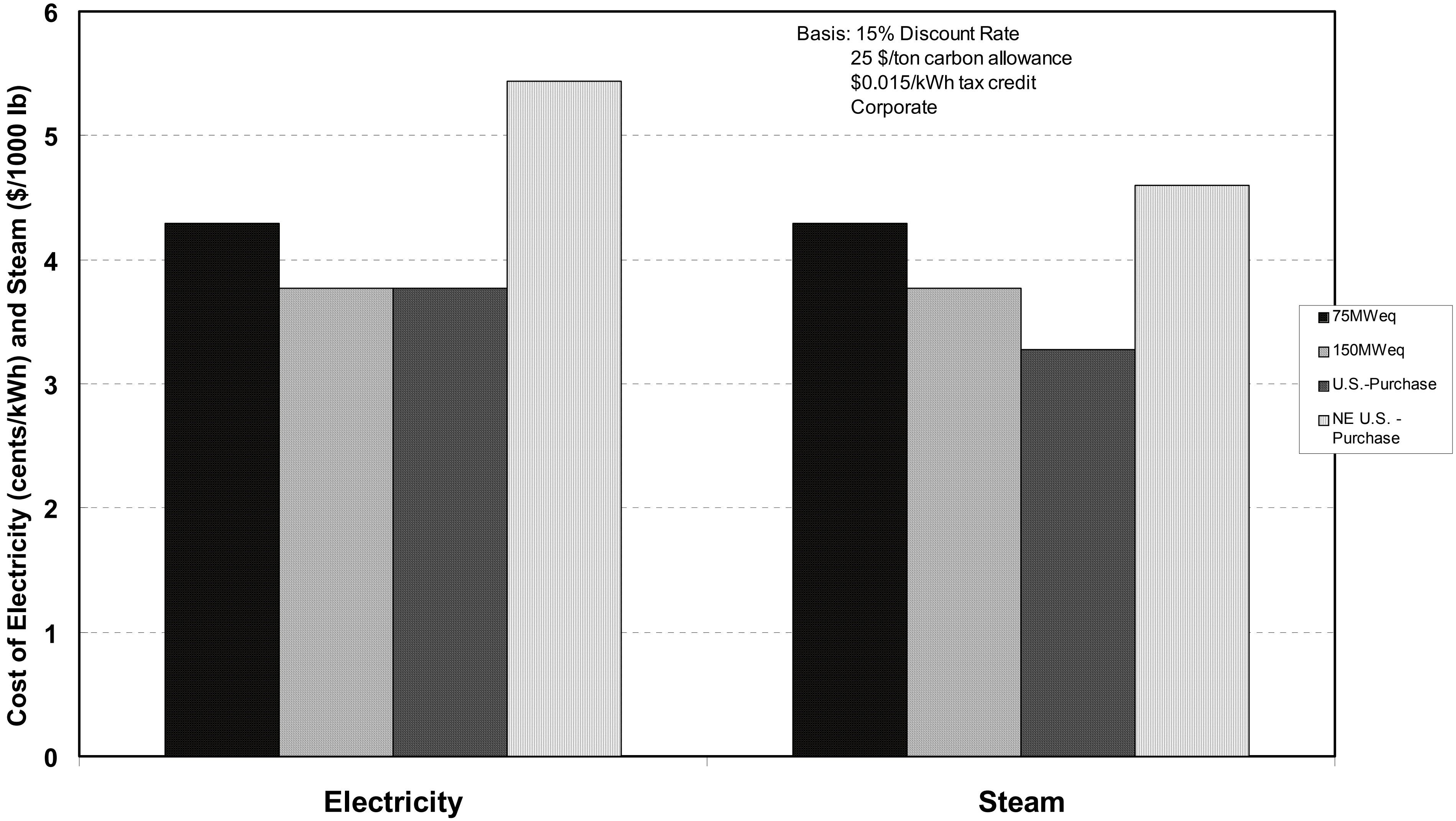


Table 5.15: Cost and Required Cash Flow Summary

CHP Case	Capital Cost		Operating Cost[a]		Cumulative Required Cash Flow
	Million $	$/kW	Million $/yr	¢/kWh	Million $
Cofiring - 105 MW	16.4	156	(2.02)[b]	(0.23)[b]	72
Direct Combustion - 75 MW	131.0	1,747	10.22	1.73	479
Direct Combustion - 100 MW	160.5	1,605	13.49	1.71	593
IGCC - 75 MW	149.3	2,070	6.71	1.14	433
IGCC - 150 MW	196.7	1,312	11.75	0.99	767

(a) incremental cost
(b)exclusive of feed

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